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Simplified Injection of Oxygen Gas into an Activated Sludge Process

Las Virgenes Municipal Water District
Calabasas, CA

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SIMPLIFIED INJECTION OF OXYGEN GAS INTO AN ACTIVATED SLUDGE PROCESS

by

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16. ABSTRACT

The Las Virgenes Municipal Water District conducted a pilot investigation of the Simplox process at their Tapia Water Reclamation Facility in Calabasas, California. The Simplox process, developed by the Cosmodyne Division of Cordon International, involves covering an activated sludge aeration basin with an inflated dome and injecting high purity oxygen into the mixed liquor through a conventional coarse bubble diffuser.

The purpose of the study was to determine the operational and economic advantages, if any, of this process over a conventional activated sludge system.

The results of the study indicated that although the Simplox process produced excellent quality effluent throughout the project by most standards, it was considered to be not economically attractive to the District due, primarily, to the inability of the process to meet strict local effluent discharge requirements of 5 mg/l each for BOD and suspended solids and the need for constant repairs to the inflated dome.

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimonies to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution; it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems to prevent, treat, and manage wastewater and solid and hazardous waste pollutant discharges from municipal and community sources; to preserve and treat public drinking water supplies; and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research and provides a most vital communications link between the researcher and the user community.

This report describes a feasibility study of a novel,covered tank, oxygen activated sludge treatment concept. An inflated polyvinyl dome was utilized to prevent the enriched-oxygen off gas from escaping to the atmosphere. Existing air aeration equipment was employed to recirculate the off gas through the mixed liquor. The information documented in this report should be of interest to design engineers and municipal officials who are considering techniques for converting air activated sludge systems to oxygen activated sludge systems.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

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ABSTRACT

The concept of using a single-stage, covered, activated sludge aeration basin and conventional aeration equipment for high purity oxygen injection was devised in 1970 by the Cosmodyne Division of Cordon International. The concept was termed the Simplox process and is investigated in this report. The Las Virgenes Municipal Water District conducted this study at their Tapia Water Reclamation Facility in Calabasas, California. The objectives of the study were:

1. to determine the practicality and cost of using this simplified method of high purity oxygen injection as an upgrading tool.

2. to determine if aeration basin throughput could be at least doubled,

3. to determine the extent of waste sludge volume reduction, and

4. to compare the results with data collected using conventional air aeration.

Conversion of the existing facilities involved covering an aeration basin with a heavy-duty plastic cover. The air blower system was modified to draw atmosphere from under the plastic cover and recirculate it thorugh the aeration basin mixed liquor using the existing coarse bubble air diffuser system. Pure oxygen from a tanker truck was vaporized and fed to the mixed liquor through a new fine bubble diffuser system (Saran wrapped tubes), which was installed at the head end of the aeration basin. A portion of the atmosphere under the plastic cover was bled off to remove carbon dioxide. Primary effluent was added to the aeration basin using step feed. Mixed liquor was clarified using conventional secondary clarifiers from which settled sludge was withdrawn for recycling and wasting.

The following conclusions were reached from this study:

1. The oxygen dispersing equipment required more operating time than conventional air compressors.

2. The process performed best at higher BOD loadings.

3. Nitrification was easily accomplished.

4. Low detention times, 1.7 hrs, produced nearly 90 percent BOD removal.

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5. Relatively high SVI values were produced.

6. Solids production was lower than in a conventional activated sludge process.

7. The use of a polyvinyl basin cover proved to be unsatisfactory because of leakage.

8. The process did not produce an effluent capable of meeting the District's strict local discharge requirements of 5 mg/l BOD and 5 mg/l suspended solids.

Based primarily on conclusions 1, 5, 7, and 8 above, the use of the Simplox process as a full-plant conversion was not considered to be economically attractive for the District.

This report was submitted in fulfillment of Grant No. S802356 by the Las Virgenes Municipal Water District under partial sponsorship of the U.S. Environmental Protection Agency and covers the period of July 1971 through February 1974.

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The research consultants, Dr. Andrew Gram and Mr. Milton Spiegel, provided valuable guidance on all phases of the project.

The personnel at the Tapia Water Reclamation Facility should be acknowledged for their analytical and operational work. The key persons were:

Mr. William Ruff, Senior Lab Technician
Mr. Roy Gull, Plant Foreman
Mr. Robert Hensley, Lab Technician

Mr. William McCumber of Cosmodyne Corporation provided instrumentation and installation services at the start of the project.

SECTION 1

INTRODUCTION

This report discusses a study which investigated the use of high purity oxygen gas as a simplified upgrading technique for activated sludge processes. This simplified technique utilized a single-stage, covered air aeration basin and minor modifications to existing blower and air injection equipment in an attempt to minimize the expense of upgrading an activated sludge process with oxygen. The objectives of the study were:

1. to determine the practicality and cost of using this simplified method of high purity oxygen injection as an upgrading tool,

2. to determine if aeration basin throughput rate could be at least doubled,

3. to determine the extent of reduction of waste activated sludge production, and

4. to compare results with baseline data collected on a parallel conventional air aeration system.

The Las Virgenes Municipal Water District (LVMWD) has constructed approximately one-fifth of the total treatment capacity necessary for ultimate growth. The District's prime objective in participating in the study was to ascertain whether savings could be realized in future construction costs (reduced tank capacity) while still achieving a superb quality effluent.

SECTION 2

CONCLUSIONS

The following conclusions were reached from this study:

1. Operation procedures for the Simplox system did not require any new operation techniques that would be unfamiliar to normal activated sludge operation. However, operation of the oxygen dispersing equipment did require more time than the operation of a standard air compressor utilized in an air activated sludge system.

2. The Simplox system performed best at the higher BOD loadings. This coupled with the relative ease and low cost for conversion of an existing air activated sludge system to an oxygen enriched system could make the Simplox system a viable system for facilities with seasonal high BOD waste loads.

3. Nitrification was easily accomplished when mixed liquor dissolved oxygen concentrations were high enough to maintain dissolved oxygen in the sludge blanket of the secondary clarifiers.

4. Aeration detention times as low as 1.7 hrs were studied with BOD removals still approaching 90 percent. Even at 1.7 hrs and the low dissolved oxygen (D.O.) conditions observed in the mixed liquor, nitrification was 83 percent complete.

5. The process produced relatively high sludge volume index (SVI) values; however, high SVI values will not necessarily reflect poor effluent quality, as long as secondary clarifier loadings are somewhat conservative.

6. The amount of excess solids production per pound of substrate removed was less than in the conventional activated sludge process. This is because the "Simplox" process provides a greater oxygen transfer rate to the mixed liquor, allowing a higher mixed liquor volatile suspended solids (MLVSS) concentration.

7. Use of a polyvinyl tent to cover the aeration basin is totally unsatisfactory. Leaks in the tent wasted quantities of oxygen and required constant mending and remending.

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8. Oxygen consumption calculations are not felt to be completely accurate for any of the phases. Discrepancies in correcting for leaks in the tent are the cause of the discrepancies.

9. Although the Simplox process produced an excellent effluent quality based on nationwide standards, the process could not produce an effluent quality satisfactory to Las Virgenes Municipal Water District needs, i.e., BOD and suspended solids concentrations of 5 mg/l each.

10. Use of the Simplox process is not economically attractive for the Las Virgenes Municipal Water District from an operation and maintenance viewpoint.

SECTION 3

BACKGROUND

The concept of using a single-stage, covered/aeration basin and conventional air aeration equipment for high purity oxygen injection was devised in 1970 by the Cosmodyne Division of Cordon International and given the name Simplox and is schematically illustrated in Figure 1. In April 1971, the Las Virgenes Municipal Water District applied to the United States Environmental Protection Agency (EPA) for a matching grant to investigate the Simplox concept.

In June 1971, EPA awarded a grant of $186,000 to the LVMWD. In addition to the EPA grant, LVMWD contributed $62,000 for a total project cost of $248,000. The study was conducted at the LVMWD Tapia Water Reclamation Facility, Calabasas, California, which had sufficient capacity for the study due to a recent expansion.

Conversion of existing facilities involved covering a existing aeration basin with a heavy duty plastic cover (the tent). The existing air blower system was modified to draw atmosphere from under the tent and recirculate it through the aeration basin mixed liquor using the existing coarse bubble air diffuser system. Liquid oxygen from a tanker truck was vaporized and fed to the mixed liquor thorugh a new fine bubble diffuser system (Saran wrapped tubes), which was installed at the head end of the aeration basin. A portion of the atmosphere under the tent was bled off to remove carbon dioxide produced by metabolic stabilization of substrate. Primary effluent was added to the aeration basin using the step feed flow reqime. Mixed liquor was clarified using a conventional secondary clarifier(s) from which settled sludge was withdrawn for recycling and wasting.

THEORY OF PROBLEM SOLUTION

Maintenance of D.O. in the mixed liquor of the activated sludge process is essential for growth of aerobic organisms. It is generally agreed, however, that D.O. concentrations above a certain critical concentration have no effect upon the rate of substrate stabilization. Various researchers have reported this critical D.O. concentration to be in the range of 0.1 - 0.5 mg/l, the discrepancies probably occurring since it is actually the D.O. concentration within the microbial floc which is critical, rather than the gross mixed liquor D.O. concentration.

Transfer of oxygen from the gas phase to the liquid phase has traditionally been accomplished in the activated sludge process by coarse bubble aeration, which in addition to transferring oxygen to the liquid phase also

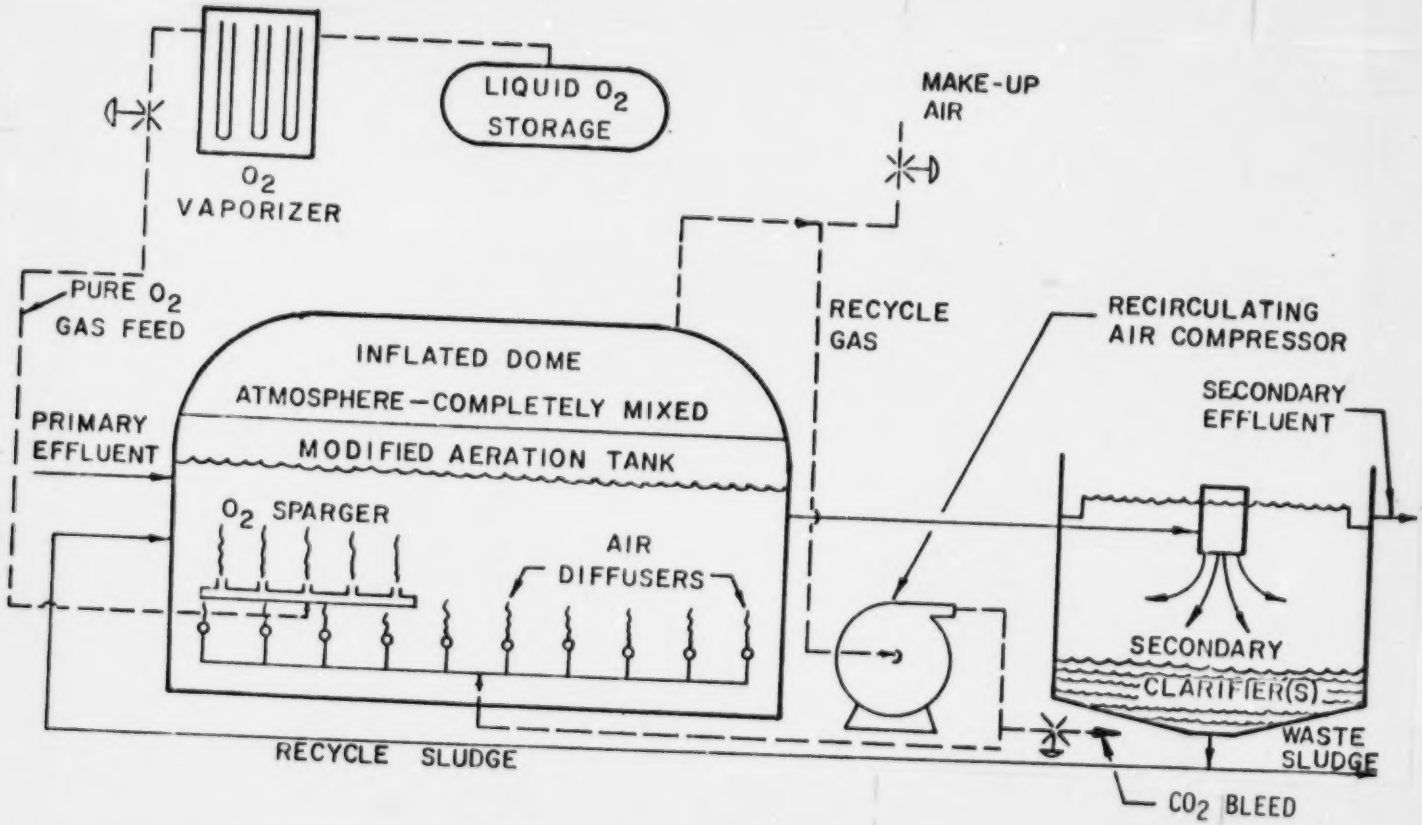

Figure 1. Schematic diagram of the Simplox process.

provides mixing energy to prevent solids deposition and scrubs metabolically produced carbon dioxide from the mixed liquor. According to Henry's Law, the D.O. saturation concentration in the mixed liquor is proportional to the oxygen concentration in the gaseous phase. Therefore, use of a gaseous phase with a higher oxygen concentration than if air aeration is used. More importantly, however, the rate of mass oxygen transfer between the gaseous and liquid phases is proportional to the difference in concentration between saturation D.O. concentration and operating D.O. concentration. For example, when the mixed liquor D.O. is 6 mg/l, the driving forces of air and a 50 percent oxygen atmosphere are as follows:

System	Saturation Conc. @ 20°C	Driving Force
Air	9.3 mg/l	9.3-6.0 = 3.3 mg/l
50% O_2 atm	22.3 mg/l	22.3-6.0 = 16.3 mg/l

Therefore, aeration using a 50 percent oxygen atmosphere system will transfer oxygen to the liquid phase at a rate of 16.3/3.3, or 4.94, times as fast as aeration with air.

The significance of using a higher oxygen concentration atmosphere for aeration is, therefore, not that a higher dissolved oxygen could be attained per se, but rather that a higher oxygen mass transfer rate is possible from the gas to the liquid phase. A higher transfer rate will allow a higher microbial concentration to be maintained in the aeration basin, and therefore, a smaller aeration basin is required to maintain the same food-to-micro-organism (F/M) loading.

In situations where an existing activated sludge plant is organically overloaded, use of the Simplox process will provide this higher oxygen mass transfer rate, thus allowing a higher MLVSS concentration to be maintained and allow the F/M loading to be lowered to a value which will produce better sludge settling in the secondary clarifier.

The project was initially planned to be divided into a five-phase study. Basically, each phase was intended to provide successive decreases in aeration detention time and successive increases in secondary clarification overflow rates.

Decreasing detention time was to be achieved through a combination of increased flow to the aeration basin and partitioning off of a portion of the aeration basin to decrease its volume. Increasing secondary clarifier overflow rates were obtained by increasing flow, but were restricted from going too high by supplementing the rectangular secondary clarifier with a circular secondary clarifier.

Due to a number of unforeseen circumstances, deviations were required from the planned experimental program. Principal among these circumstances were a lack of available primary effluent step feed ports in the tent, equipment malfunctions, and bulking in the secondary sedimentation basin(s).

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Table 1 presents a listing of phases as they were actually conducted and as they are designated in this report. A more detailed discussion of operational problems will be included with the discussion of results for each individual phase.

TABLE 1. EXPERIMENTAL PROGRAM

	Phase 1	Phase 2	Phase 3	Phase 4	Phase 5	Phase 6
Influent flow, m^3/day (mgd)	3785 (1.000)	7116 (1.880)	3838 (1.014)	4940 (1.305)	5545 (1.465)	7014 (1.853)
Amount of aeration basin used, %	100	100	46	46	46	46
Aeration basin detention time ($Q+Q_R$), hr	7.20	3.53	3.24	2.39	2.14	1.70
Number of secondary clarifiers used	1	1	2	2	2	2
Clarifier overflow rate; Rectangular, m^3/day/m^2 (gpd/ft^2)	17.0 (417)	22.1 (543)	14.0 (343)	11.0 (270)	12.1 (297)	16.4 (402)
Circular, m^3/day/m^2 (gpd/ft^2)			13.4 (330)	16.8 (412)	19.2 (471)	22.4 (550)

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SECTION 4

PHYSICAL DESCRIPTION

EQUIPMENT FABRICATION

As a result of a recent expansion of the Tapia Water Reclamation Facility to 31,280 m^3/day (8 mgd), a portion of the original plant was available for use in this study. Primary sedimentation basin effluent was obtained from the new plant addition.

Aeration Basin

The aeration basin had dimensions of 35.4 m long x 9.14 m wide x 4.57 m deep (116.25 ft x 30 ft x 15 ft) and an effective volume of 1,481,300 l (391,361 gal). The entire aeration basin was used for Phases 1 and 2. Prior to beginning Phase 3, a partitioning baffle was placed in the aeration basin to decrease its effective volume. Placement of this baffle decreased the aeration basin length to 16.46 m (54 ft) resulting in an effective volume of 687,980 l (181, 764 gal). Addition of primary effluent to the aeration basin was by step feed. Primary effluent was added to the aeration basin at three locations in Phases 1 and 2 at the head end, 9.14 m (30 ft) down the aeration basin, and 18.3 m (60 ft) down the aeration basin. In Phases 3, 4, 5, and 6, primary effluent was only introduced at the first two of these locations. Return activated sludge was introduced at the head end of the aeration basin in all phases (Figure 2).

The aeration basins were covered by a tent to allow maintenance of an oxygen concentration of about 50 percent (Figure 3). The tent was constructed from polyethylene. The tent was attached to the aeration basin by wrapping it under sections of lumber, which were in turn bolted to the concrete at the outer surface of the aeration basin. The tent had a door fabricated in it to allow the entry of personnel into the tent for operation and maintenance purposes. Throughout all phases of the study, leaks were routinely noted in the tent, either visually or by smoke detection. These leaks were attributed to normal wear of the fabric, stress caused by wind, and bullet holes caused by pranksters. As holes were detected, they were sealed using glue and polysheets.

Atmosphere from under the tent was recirculated through the existing air headers in the aeration basin. A new 849.4 l/sec (1,800 cfm) blower was equipped to perform this function (Figure 4). Suction for the blower was taken from under the tent near the effluent end of the aeration basin. The rate of atmospheric recirculation was established to provide both sufficient mixing and a mixed liquor D.O. concentration above the critical concentration.

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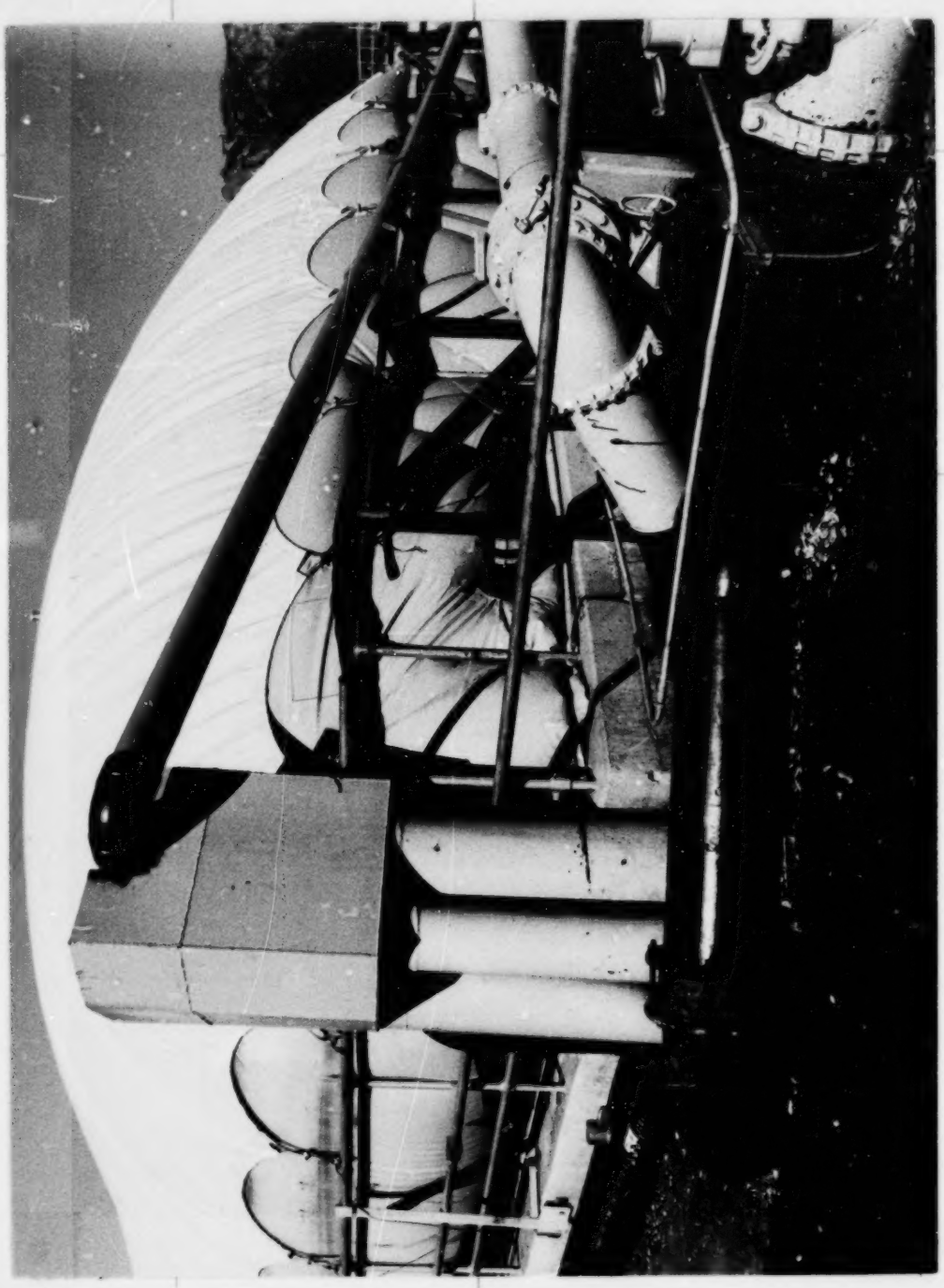

Figure 2. RAS and WAS diversion box.

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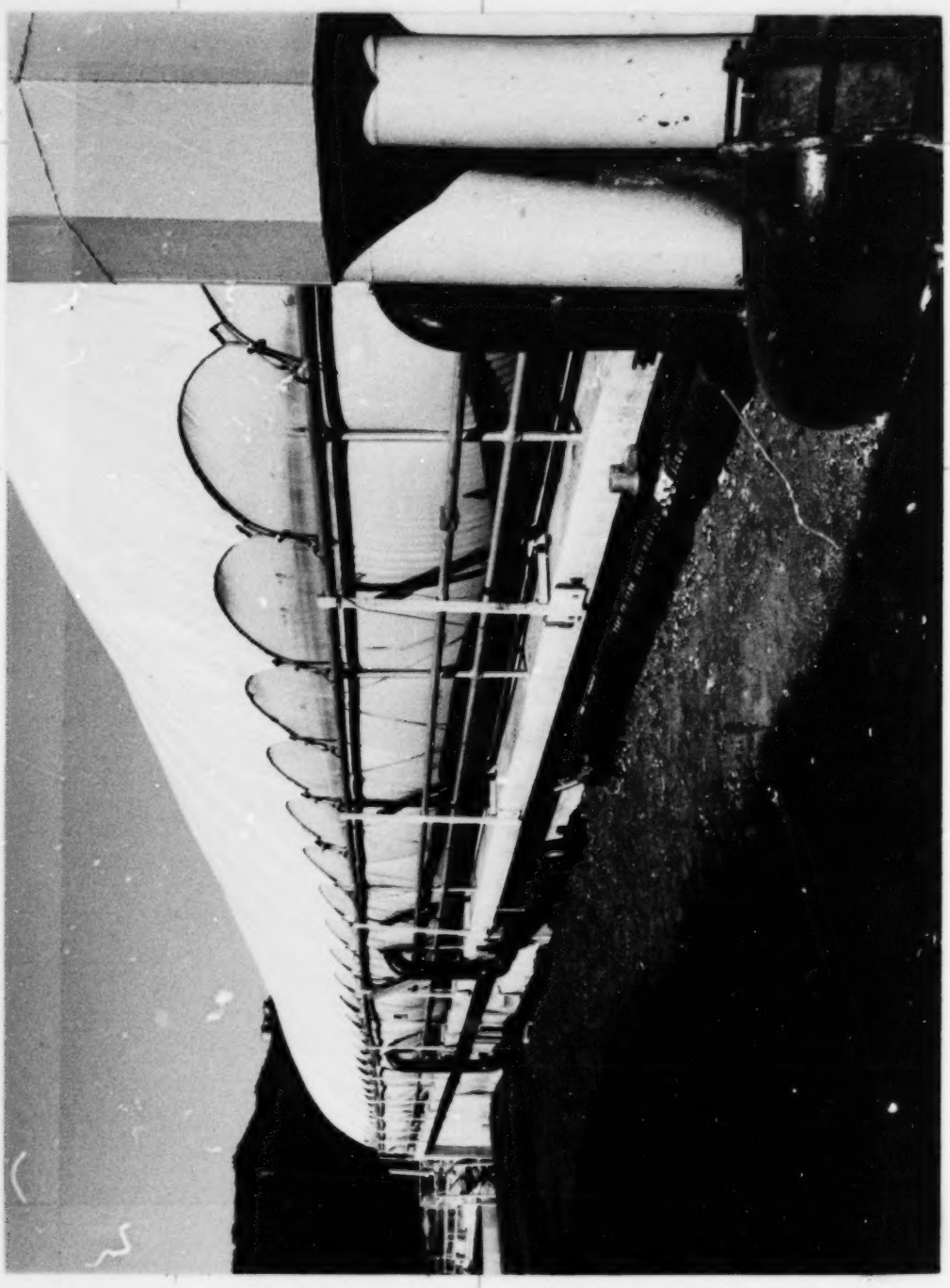

Figure 3. Aeration tank tent and step feed piping.

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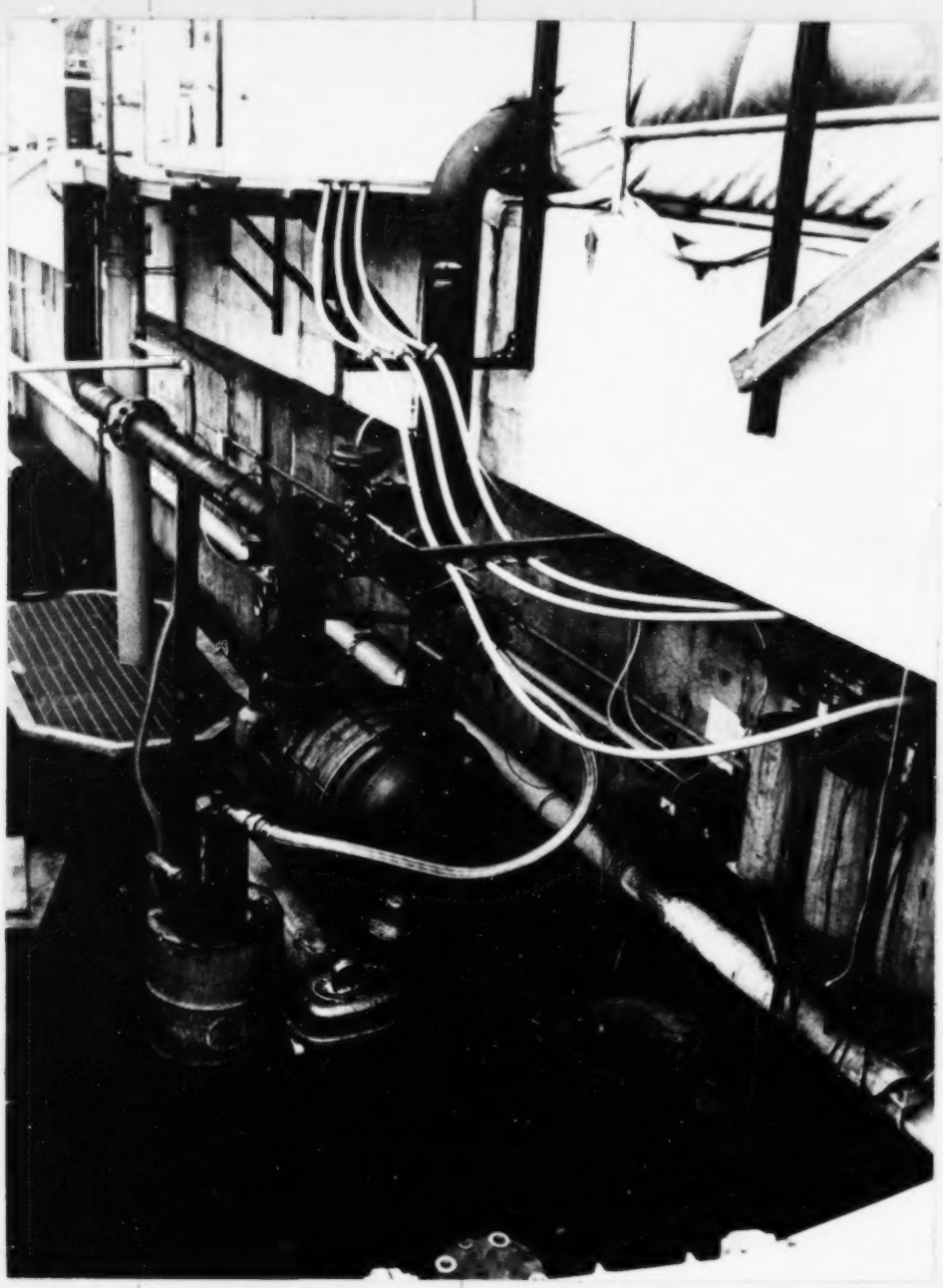

Figure 4. Aeration mixing blower, rotometer, and sample ports.

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During Phase 1, recirculation was at a rate of 307.2 l/sec (651 cfm) with a mixed liquor D.O. of about 8.5 mg/l. Recirculation rate followed a downward trend as the study progressed, ending with a rate of 109 l/sec (231 cfm) and a mixed liquor D.O. of about 1.0 mg/l in Phase 6. Sufficient mixing was provided even at the lower flow rate.

In addition to the existing air header system used for atmosphere recirculation, a separate system was installed for addition of pure oxygen to the system. This system consisted of a submerged header, 12.19 m (40 ft) long, at the end of the aeration basin and on the opposite side from the aeration diffusers. On the header, Saran wrapped tubes were installed for fine bubble diffusion of the pure O_2 into the mixed liquor.

Secondary Clarification

Two secondary clarifiers were used in this study. A rectangular secondary clarifier with dimensions 36.58 m long x 6.10 m wide x 3.05 m average surface water depth (120 ft x 20 ft x 10 ft) was used in all phases of the study. The surface area of this clarifier was 223 m^2 (2,400 ft^2), and total weir length was 40.54 m (133 ft). The rectangular clarifier was equipped with a mechanical surface skimming device. Sludge was withdrawn from the rectangular clarifier using variable speed, centrifugal pumps with a maximum capacity of 25.1 l/sec (400 gpm) each.

A circular secondary clarifier was also used in parallel operation with the rectangular secondary clarifier during a portion of Phase 2 and during all of Phases 3, 4, 5, and 6. The circular clarifier was 13.72 m (45 ft) in diameter and had an average surface water depth of 3.05 m (10 ft). The total surface area was 148.2 m^2 (1,595 ft^2), and total weir length was 43.07 m (141.3 ft). The circular clarifier was equipped with a mechanical surface skimming device. Sludge was withdrawn using a variable speed, centrifugal, underflow pump, capable of operating at a pumping rate up to 25.2 l/sec (400 gpm).

Parallel operation of the rectangular and circular secondary clarifier is referred to as the combined secondary sedimentation system.

Pure Oxygen Storage and Delivery System

Pure oxygen was delivered to the Tapia facility in a 1.5 m^3 (4,000 gal) tank truck by Cosmodyne (Figure 5). Liquid O_2 was transferred from the delivery trucker to the O_2 storage tanker. Pure oxygen in liquid form was withdrawn from the tanker and passed through an oxygen vaporizer at ambient temperatures (Figure 6).

Instrument and Control System

In an effort to achieve optimum results from a minimum expenditure, spare components were used in the existing Tapia plant instrumentation system.

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Figure 5. Liquid oxygen tanker.

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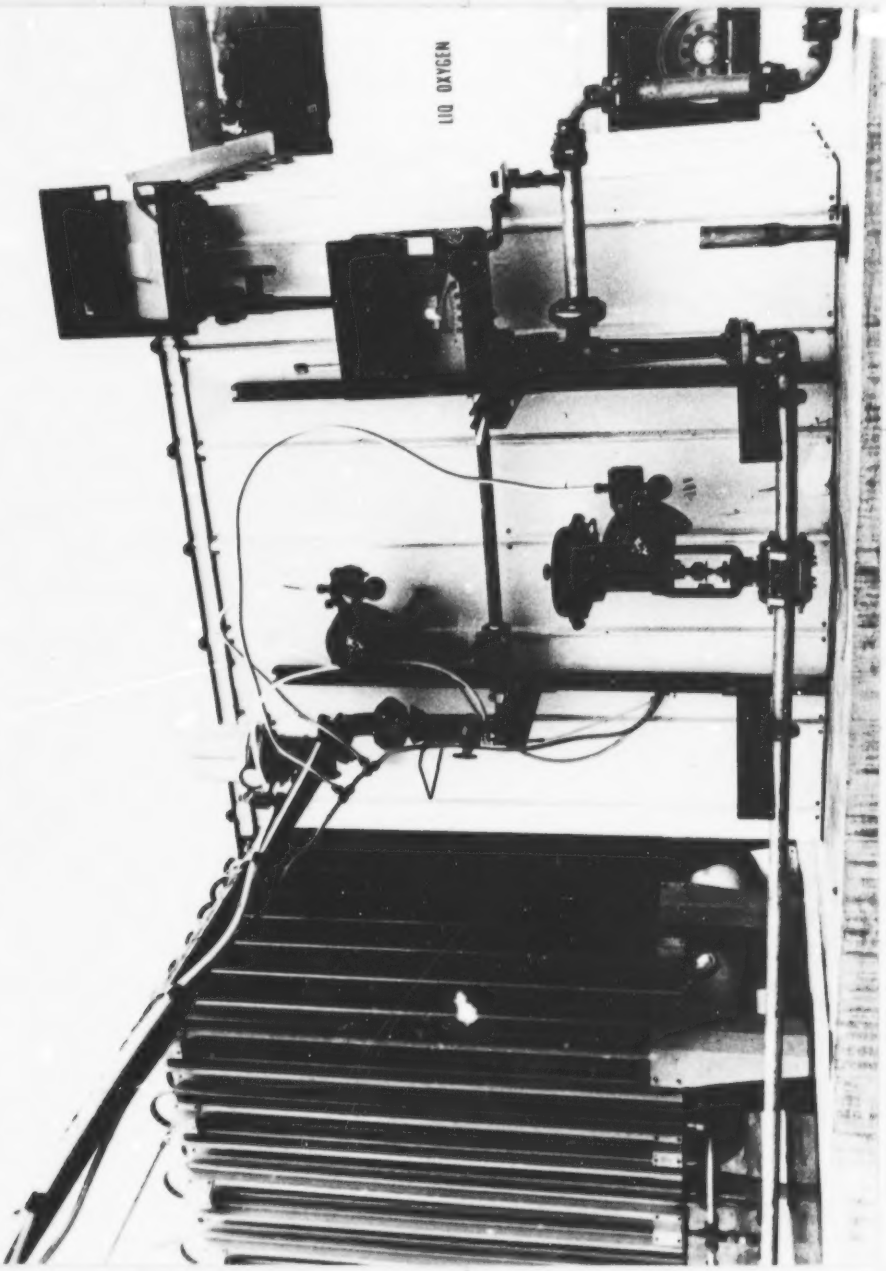

Figure 6. Oxygen vaporizer and gas feed control system.

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A flow diagram was developed, and points requiring instruments and controls were established. A comprehensive control system was applied to this composite using instruments compatible with existing new system. Instrumentation and controls are divided into numbered loops for ease of identification. Instrumentation loops 1 through 8 are metering only, using propellor type meters. Only loop 1, influent flow, is recorded and totalized on the main control panel.

Loop 9 is probably the most complex, metering and controlli..g oxygen flow to the oxygen diffuser. Using cascaded feed-forward control, the primary variable is sensed by the oxygen analyzer as percentage of oxygen in the atmosphere under the tank cover. A controller with proportional plus reset modes using primary variable as process input, has its output cascaded into the set point of the oxygen flow controller which uses the secondary variable, oxygen flow, as process input. Thus the oxygen flow control is a closed loop, the set-point of which is continually adjusted by the oxygen analysis controller to maintain the set percentage of oxygen in the atmosphere under the cover.

Loop 10 controls the recirculation flow of the under-cover gas through the blower and the diffusers to provide tank agitation and additional oxygen uptake. Recirculation gas flow is sensed as ΔP across an orifice plat and controlled by a butterfuly valve in the blower suction line. A manual bypass is provided around the blower to maintain blower air flow above the surge point if low recirculation flows are desired. Loops 11 through 16 are flow elements in the recirculation gas line to provide balancing flow indication to adjust flow to individual diffusers.

Loop 17 senses structure pressure and bleeds off excess undercover gas to maintain structure support pressure.

Using the new plant spare instrumentation facilities and main instrument panel all recorded variables were electronically transmitted and recorded on the main panel.

Electronic controllers are also located on the main panel so that the process variables may be monitored and set from the main control room.

Control valves were pneumatically operated from electronic input signals using current to air transducers.

Sewage and sludge flows are locally metered and manually controlled.

During the course of the test, difficulties were encountered which prevented the instrumentation and control system from operating as planned and installed.

The major difficulty was encountered in leakage of the gas under the tank cover to the outside atmosphere. Because of leakage, between the concrete wall and the fabric structure at this pressure, it was impossible to maintain adequate support for the structure without adding atmospheric air to the closed system. This was accomplished by bleeding air from the aeration air

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serving an adjacent tank into the suction of the blower for the test tank. This air quantity was measured by a large positive displacement gas meter initially and by a rotometer after the gas meter failed in service.

The leakage also prevented the cascaded feed-forward control of the oxygen content of the atmosphere under the cover, so the oxygen feed was manually adjusted using flow controller FIC-9. As surplus gas was never encountered, bleed to the atmosphere through loop 17 was not needed except as a safety backup and FCV-17 remained closed thorughout the test.

The propellor meters used for determining influent and RAS flows worked well on influent but were difficult to keep in operation in sludge service.

Because of continuing clogging and loss of suction in syphoning sample lines to the mixed liquor O_2 analyzers, redesign and installation of purge units was necessary. Once this was accomplished, continuous monitoring records were fairly dependable. O_2 analyzers A-18, A-19, and A-20 showed D.O. levels to be equal throughout the length of the tank.

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SECTION 5

OPERATION AND EVALUATION PHASE

The study was initiated on April 6, 1972 with a brief shakedown period, during which instruments were checked out, calibrated, and adjusted. Mixed liquor solids were built up by the addition of waste activated sludge (WAS) from the main plant. Within a few days, mixed liquor solids reached a sufficiently high concentration, and no further WAS from the main plant was introduced. By April 17, 1972 effluent quality had stabilized at a BOD of about 15 mg/l and SS of about 10 mg/l at a primary effluent flow of 0.0307 m^3/sec (0.7 mgd). Pure oxygen was first added to the aeration basin on April 17, 1972 and after a week long acclimation period, Phase 1 was begun.

As previously discussed, initial plans were to conduct the study in five separate phases, but deviations in the planned study were necessary due to a number of operational and equipment problems. In all, six separate phases were conducted as shown in Table 1. Each of the six phases is discussed separately in the following sections.

PHASE 1

Background

Phase 1 began on April 25, 1972 and was completed on July 31, 1972, a period of 98 days. Phase 1 was planned to be an ultra-conservative phase which would allow comparison with the latter stages of the study. Flow to the aeration basin during Phase 1 was 0.0438 m^3/sec (1 mgd), with an average return activated sludge rate of 30 percent. The flow was continuous, with no diurnal variation, and was taken directly from the primary effluent stream going to the main air-activated sludge process. Aeration basin detention time was 7.2 hrs and mean cell residence time (MCRT) was 108 days. No solids were wasted from the system except secondary clarifier skimmings and suspended solids in the secondary effluent.

Operational Problems

The major operational problems encountered in Phase 1 were bearing problems on the recirculation blower, leaks in the tent, and failures of the aeration basin D.O. analyzer sampling pots.

The blower problems resulted from a motor shaft which was initially defective and caused rapid wear of the shaft bearings. The blower was shut down for maintenance several times during Phase 1, usually for only an hour or two, but once for a 2-day period (May 30 and 31). Pure oxygen was added continuously to the aeration basin, even during periods when recirculation blower was down.

As previously mentioned, leaks in the tent were a persistent problem throughout the entire study. In order to maintain a positive pressure in the tent, air was bled into the atmosphere under the tent at an average rate of 3.54 l/sec (7.5 cfm). Numerous smoke tests were conducted throughout Phase 1 to locate leaks in the tent.

Aeration Basin-Secondary Clarifier Efficiency

Overall system efficiency can be best judged by removals through the aeration basin secondary clarifier complex. As shown in Table 2, removals were exceptionally good for organic substrates, suspended solids, and ammonia nitrogen and total Kjeldahl nitrogen (TKN). Almost complete nitrification occurred, as shown by the 97 percent reduction of ammonia nitrogen. A pH decrease also occurred, due to removal of alkalinity.

As a result of the low substrate concentrations in the primary effluent and the large volume of the aeration basin, F/M loadings were very low, as may be seen in Table 3.

For example, the F/M loading was only 0.073 mass BOD applied/day/mass MLVSS. These low loadings resulted in a slightly high average SVI, 115 ml/gm, but did not adversely affect substrate removal efficiency, as seen in Table 2. Probably a key factor in the excellent substrate removal was the low secondary clarifier loading rate and rapid removal of settled sludge for recycle, as well as the low basin and weir overflow rates in the secondary clarifier, as shown in Table 3.

TABLE 2. PHASE 1 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	82	2	97.6
T-COD, mg/l	153	35	77.1
S-COD, mg/l	58	16	72.4
Suspended solids, mg/l	73	9	87.7
Volatile suspended solids, mg/l	72	7	90.3
NH_3-N, mg/l	13	0.4	96.9
NO_3-N, mg/l		16.2	
Total TKN, mg/l		1.2	
Alkalinity, mg/l	242	145	40.1
Turbidity, JTU		1.4	
pH, units	7.5	6.6	

TABLE 3. PHASE 1 LOADING PARAMETERS

F/M Ratios	
mass BOD applied/day/mass MLVSS	0.073
mass T-COD applied/day/mass MLVSS	0.133
mass S-COD applied/day/mass MLVSS	0.053
Volumetric Loadings	
kg BOD applied/day/m^3	0.215
(lb BOD applied/day/100 ft^3)	(13.43)
kg T-COD applied/day/m^3	0.392
(lb T-COD applied/day/1000 ft^3)	(24.48)
hg S-COD applied/day/m^3	0.156
(lb S-COD applied/day/100 ft^3)	(9.75)
Substrate Removals	
mass BOD removed/day/mass MLVSS	0.070
mass T-COD removed/day/mass MLVSS	0.102
mass S-COD removed/day/mass MLVSS	0.036
Rectangular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	25.14
(gpd/ft^2)	(617)
Weir overflow rate, m^3/day/m	93.39
(gpd/ft)	(7519)
Solids loading rate, kg MLSS/day/m^2	62.15
(lb MLSS/day/ft^2)	(12.73)
Circular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	-
(gpd/ft^2)	-
Weir overflow rate, m^3/day/m	-
(gpd/ft)	-
Solids loading rate, kg MLSS/day/m^2	-
(lb MLSS/day/ft^2)	-

Sludge Characteristics

The low loading resulted in an extremely long MCRT of 108 days, in spite of a MLVSS of 2,950 mg/l. As a result of the long MCRT, the sludge growth was well into the endogenous growth phase. This is also evidenced by the low sludge production rates, as shown in Table 4, and the absence of any direct wasting of activated sludge. The only solids wasted from the system were secondary clarifier skimmings, which averaged 32.65 kg (72 lb) VSS/day, and effluent suspended solids, which averaged 26.48 kg (58.4 lb) VSS/day. The long MCRT did not significantly affect the return activated sludge (RAS) concentration, which averaged 14,324 mg/l and 82.7 percent volatile content.

Oxygen Utilization and Nitrogen Conversion

Oxygen utilization efficiency (Table 5) was relatively high, averaging 85.2 percent, as was the mixed liquor D.O. which averaged 7.5 mg/l. The amount of oxygen consumed per quantity of substrate stabilized was extremely high, even after correcting for oxygen required for biological oxidation of ammonia by nitrifying bacteria. The explanation for the high oxygen consumption values is most likely the large amount of oxygen required for endogenous respiration of cellular material. This theory is enhanced by considering that only 59.13 kg (130.4 lb) of volatile solids were wasted per day, while 302.58 kg (667.2 lb) of BOD_5 were oxidized per day.

TABLE 4. PHASE 1 SLUDGE CHARACTERISTICS

MLSS, mg/l	3,692
MLVSS, mg/l	2,950
Return Activated Sludge TSS, mg/l	14,324
Return Activated Sludge Rate, % Q	
Rectangular clarifier	30
Circular clarifier	-
SVI, ml/gm	115
Mean Cell Residence Time, days	108
Sludge Production	
mass VSS produced/mass BOD removed	0.195
mass VSS produced/mass T-COD removed	0.132
mass VSS produced/mass S-COD removed	0.372

TABLE 5. PHASE 1 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen supplied,	
kg O_2 supplied/m^3	0.327
(lb O_2 supplied/10^6 gal)	(2733)
mass O_2 supplied/mass BOD removed	4.10
mass O_2 supplied/mass T-COD removed	3.16
Oxygen consumption,	
kg O_2 consumed/m^3	0.279
(lb O_2 consumed/10^6 gal)	(2329)
mass O_2 consumed/mass BOD removed	3.50
mass O_2 consumed/mass T-COD removed	2.70
Oxygen utilization efficiency, %	85.2
Average aeration basin DO, mg/l	7.5
Ammonia oxidation, %	97
Rate of nitrification, mass NH_3-N oxidized/day/mass MLVSS	0.0203
Alkalinity consumption, mass alkalinity consumed/mass NH_3-N oxidized	7.7

Almost complete nitrification of ammonia, 97 percent, occurred in Phase 1. This could be anticipated, however, because the MCRT of 108 days exceeds the approximate 7-day MCRT required for nitrification, and rapid sludge removal in the secondary clarifier prevented anaerobisis of the sludge.

Conclusions

High purity oxygen aeration with an aeration time of 7.2 hrs and a mean cell residence time of 108 days provides both exceptionally good effluent quality and exceptionally low solids wasting requirements. These results are expected, however, because MLVSS are obviously maintained well into the endogenous growth phase at such a high mean cell residence time.

PHASE 2

Background

Phase 2 began on September 11, 1972 and was terminated on November 13, 1972, a period of 64 days. The period between Phases 1 and 2 was necessary

for factory repair of the recirculation blower motor and buildup and acclimation of MLVSS following return of the blower motor. Because primary effluent available for the study was insufficient to allow operation at the planned flow rate, a portion of effluent from the combined secondary sedimentation system was returned to the head works section of the plant to mix with raw sewage. In order to prevent a substantial dilution of substrate concentration by return of this secondary effluent, all but one of the weirs in the existing primary clarifier were blocked off in an attempt to decrease efficiency of the primary sedimentation basin. This turned out to be an extremely difficult process to regulate because the sludge blanket depth in the primary clarifier was not far below the weirs. At times, this method of operation allowed unxepectedly large concentration of substrate and solids to pass over the one remaining weir, and thus overload the aeration basin.

Average flow to the aeration basin in Phase 2 was 7225.8 m^3/day (1.88 mgd), and the return activated sludge flow averaged 42.5 percent. Average aeration basin detention time was 3.53 hrs, and average mean cell residence time was 48.6 days.

Operational Problems

It was observed that in addition to loss of atmosphere through leaks in the tent, atmosphere was also being lost through the discharge line from the aeration basin to the secondary clarifiers. Several attempts were made to correct this condition, but they resulted in flow restriction during periods of peak flow and overflow of the aeration basin.

Problems were also encountered in maintaining a sufficient RAS pumping rate from the rectangular secondary clarifier. This resulted in periodic solids bulking in the rectangular secondary clarifier. The problem was finally solved by decreasing the flow to the rectangular clarifier, and thus, the solids loading rate. The concentration of the RAS then decreased, and the rate of pumping went up.

Another problem deserving mention is an unexplainable pH increase in the primary effluent for a week in early October. Aeration basin pH increased gradually from 6.3 to 7.3 over a 5-day period, then abruptly dropped back to 6.3 in a 1-day period.

Aeration Basin-Secondary Clarifier Efficiency

Removal efficiencies in Phase 2 (Table 6) were affected by period high substrate and solids carryover from the primary clarifier aeration basin, as a result of secondary effluent recirculation and the intentional blockage of primary clarifier weirs. This method of primary clarifier operation, while serving to increase substrate concentration, resulted in a higher than normal percentage of substrate being in the particulate form. Although BOD, SS, and VSS were essentially the same as Phase 1, COD and TKN removals were much lower. The low COD removals are felt to be due to a higher percentage of non-biodegradable COD in the primary effluent compared to effluent from a normally operated primary sedimentation basin. As in Phase 1, nearly complete nitrification occurred as evidenced by a 98.8 percent decrease in ammonia. A pH

decrease again occurred due to destruction of alkalinity during nitrification. The decreased TKN removal efficiency is also felt to be related to primary clarifier operation, as well as cellular organic nitrogen contained in microbial floc during periods when sludge in the rectangular clarifier bulked.

TABLE 6. PHASE 2 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	143.5	3.7	97.4
T-COD, mg/l	145.8	94.9	34.9
S-COD, mg/l	43.0	20.1	53.3
Suspended solids, mg/l	214	13.4	93.7
Volatile suspended solids, mg/l	200	9.5	95.3
NH_3-N, mg/l	6.7	0.08	98.8
NO_3-N, mg/l		13.7	
Total TKN, mg/l	12.6	2.7	78.6
Alkalinity, mg/l	183	126	31.1
Turbidity, JTU		2.8	
pH, units	7.2	6.6	

F/M loading (Table 7) increased in response to the increase in primary effluent substrate concentration. In Phase 2, they were nearly at an "optimum level" (relative to air aeration), yet the SVI of 179 (Table 8) was noticeably higher than in Phase 1. This higher SVI, however, only reduced the RAS concentration by about 9 percent compared to Phase 1. RAS concentration was still very acceptable, averaging 13,193 mg/l.

During all of Phase 2, both the rectangular and circular clarifiers were used. As a result, secondary clarifier surface and weir overflow rates were essentially unchanged from Phase 1. Solids loading rate did increase slightly due to a higher MLSS concentration.

Sludge Characteristics

Mixed liquor volatile suspended solids were maintained at a relatively high level, 3,046 mg/l. The principal factor allowing this high a MLVSS was the relatively high volatile solids concentration of the return activated sludge, VSS = 9,400 mg/l. The ability to produce such a concentrated sludge resulted from low surface overflow rates and relatively low weir overflow rates in the combined secondary clarification system. In addition, the solids loading rate was relatively high, which could have partly offset the high SVI of 179.

MCRT decreased to 48.6 days, partially due to increased effluent VSS, but mainly due to an increase in surface skimmings from the secondary clarifier. There was no waste activated sludge from the system until the last week of Phase 2. Solids production on a BOD basis was only slightly higher than in

TABLE 7. PHASE 2 LOADING PARAMETERS

F/M Ratios:	
mass BOD applied/day/mass MLVSS	0.236
mass T-COD applied/day/mass MLVSS	0.224
mass S-COD applied/day/mass MLVSS	0.065
Volumetric loadings:	
kg BOD applied/day/m^3	0.700
(lb BOD applied/day/1000 ft^3)	(43.75)
kg T-COD applied/day/m^3	0.692
(lb T-COD applied/day/1000 ft^3)	(43.20)
kg S-COD applied/day/m^3	0.203
(lb S-COD applied/day/1000 ft^3)	(12.70)
Substrate Removals:	
mass BOD removed/day/mass MLVSS	0.230
mass T-COD removed/day/mass MVLSS	0.075
mass S-COD removed/day/mass MLVSS	0.035
Rectangular Clarifier Loadings:	
Surface overflow rate, m^3/day/m^2	17.76
(gpd/ft^2)	(436)
Weir overflow rate, m^3/day/m	97.72
(gpd/ft)	(7868)
Solids loading rate, kg MLSS/day/m^2	68.59
(lb MLSS/day/ft^2)	(14.05)
Circular Clarifier Loadings:	
Surface overflow rate, m^3/day/m^2	22.12
(gpd/ft^2)	(543)
Weir overflow rate, m^3/day/m	76.11
(gpd/ft)	(6128)
Solids loading rate, kg MLSS/day/m^2	86.26
(lb MLSS/day/ft^2)	(17.67)

TABLE 8. PHASE 2 SLUDGE CHARACTERISTICS

MLSS, mg/l	3,896
MLVSS, mg/l	3,046
Return Activated Sludge TSS, mg/l	13,193
Return Activated Sludge Rate, % Q	
Rectangular clarifier	55.0
Circular clarifier	26.8
SVI, ml/gm	179
Mean Cell Residence Time, days	48.6
Sludge Production	
mass VSS produced/mass BOD removed	0.248
mass VSS produced/mass T-COD removed	0.493
mass VSS produced/mass S-COD removed	0.896

Phase 1, which as in Phase 1, was due to the long MCRT, and, therefore, maintenance of the cells in an endogenous growth phase for substantial periods of time. Solids production on a COD basis was substantially greater than in Phase 1 due principally to a decrease in COD removal efficiency.

Oxygen Utilization and Nitrogen Conversion

As shown in Table 9, oxygen utilization efficiency was much lower than in Phase 1. This was due to an increased leakage rate through the tent and an increased amount bearing in the aeration basin effluent, a result of higher mixed liquor D.O. and higher plant flow. Oxygen consumption per pound of BOD removed dropped sharply from Phase 1. More than likely, this was due to a decrease in MCRT to 48.6, days and, therefore, a decrease in the time that the microbial cells were undergoing endogenous respiration.

Essentially, complete nitrification was achieved, as measured by a 98.8 percent reduction of ammonia nitrogen. The degree of nitrification dropped from Phase 1 because the NH_3-N concentration was only about half as much due to recycling secondary effluent to the head end of the plant.

As in Phase 1, a greater amount of alkalinity was lost from the system than stochiometrically required for ammonia oxidation. It is probable that some carbon dioxide was lost from the system due to the relatively low pH levels in the mixed liquor and secondary clarifier.

TABLE 9. PHASE 2 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen Supplied	
kg O_2 supplied/m^3	0.166
(lb O_2 supplied/10^6 gal)	(1385)
mass O_2 supplied/mass BOD removal	1.86
mass O_2 supplied/mass T-COD removed	5.30
Oxygen Consumption	
kg O_2 consumed/m^3	0.099
(lb O_2 consumed/10^6 gal)	(824)
mass O_2 consumed/mass BOD removed	1.08
mass O_2 consumed/mass T-COD removed	3.08
Oxygen Utilization Efficiency, %	59.4
Average Aeration Basin DO, mg/l	9.04
Ammonia Oxidation, %	98.8
Rate of Nitrification	
mass NH_3-N oxidized/day/mass MLVSS	0.014
Alkalinity Consumption	
mass alkalinity consumed/mass NH_3-H oxidized	9.14

PHASE 3

Background

Phase 3 began on January 22, 1973 and was terminated on April 3, 1973, a period of 72 days. Phase 3 was intended to have a shorter aeration time than Phase 2 as well as lower surface overflow rates in the secondary clarifiers. To accomplish this, a partitioning baffle was placed at approximately the midpoint of the aeration basis, thereby decreasing the volume by about half. Flow at one-half the rate in Phases 1 and 2 thus produces the same aeration detention time, for a constant recycle rate.

Average flow in Phase 3 was 3875.8 m^3/day (1.024 mgd) with an average return activated sludge rate of 33 percent. Aeration basis detention time was 3.24 hrs, and the mean cell residence time was 84 days. There was no skimming during Phase 3; solids were only removed in the waste activated sludge and in the effluent.

Operational Problems

The major operational problems encountered in Phase 3 were failure of the recirculation blower, leaks in the tent, and bulking of the final clarifiers.

As in prior phases, recurrent leaks occurred in the tent. Usually they were caused by wind, but they still indicate the drawbacks of using a plastic cover.

Bulking in the final clarifiers was attributable to several factors; density currents caused part of the problem, these could not be alleviated, however, because of the nature of clarifier design. A higher SVI than in the previous phases also influenced the frequency of bulking. The higher SVI resulted in maintenance of a1.68 to 1.98 m (5.5 to 6.5 ft) deep sludge blanket, despite a 33 percent return activated sludge recycle rate and low secondary clarifier overflow rates.

Aeration Basin-Secondary Clarifier Efficiency

Efficiency of the aeration basin-secondary clarifier complex is presented in Table 10. Substrate removals were still very high, although somewhat less than in Phase 1. The BOD concentration applied to the aeration basin was less than in Phase 2, because the primary clarifier was operated in a normal fashion, but both total and soluable COD concentration increased. Most notable is the decrease in the reduction of ammonia, averaging only 86.5 percent. A pH decrease was alkalinity by nitrifying bacteria during nitrification.

TABLE 10. PHASE 3 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	88.8	2.6	47.1
T-COD, mg/l	189	30.9	83.7
S-COD, mg/l	82.4	23.9	71.0
Suspended solids, mg/l	44.0	4.8	89.1
Volatile suspended solids, mg/l	38.4	3.6	90.6
NH_3-N, mg/l	11.83	1.6	86.5
NO_3-N, mg/l		8.0	
Total TKU, mg/l	15.6	3.0	80.8
Alkalinity, mg/l	225.8	160.2	29.1
Turbidity, JTU		1.7	
pH, units	7.43	6.74	

In Phase 3 (Table 11), the BOD F/M loading was lower than Phase 2 because primary effluent BOD decreased considerably. Conversely, total and soluble COD F/M loadings were higher than Phase 2. It is important to note that the ratio of the BOD to T-COD F/M loading in Phase 3 was 0.47, while in Phases 1

and 2 it was 0.55 and 1.05, respectively. Primary effluent substrate characteristics were, therefore, closer in Phases 1 and 2.

Surface and weir overflow rates and solids loading rate were lower than Phase 2, because total flow was less in Phase 2. These low loadings are considered very conservative.

TABLE 11. PHASE 3 LOADING PARAMETERS

F/M Ratios:	
mass BOD applied/day/mass MLVSS	0.174
mass T-COD applied/day/mass MLVSS	0.372
mass S-COD applied/day/mass MLVSS	0.161
Volumetric Loadings:	
kg BOD applied/day/m^3	0.497
(lb BOD applied/day/1000 ft^3)	(31.03)
kg T-COD applied/day/m^3	1.057
(lb T-COD applied/day/1000 ft^3)	(65.97)
kg S-COD applied/day/m^3	0.461
(lb S-COD applied/day/1000 ft^3)	(28.77)
Substrate Removals:	
mass BOD removed/day/mass MLVSS	0.163
mass T-COD removed/day/mass MLVSS	0.312
mass S-COD removed/day/mass MLVSS	0.101
Rectangular Clarifier Loadings:	
Surface overflow rate, m^3/day/m^2	13.97
(gpd/ft^2)	(343)
Weir overflow rate, m^3/day/m	76.94
(gpd/ft)	(61.95)
Solids loading rate, kg MLSS/day/m^2	53.34
(lb MLSS/day/ft^2)	(10.72)
Circular Clarifier Loadings:	
Surface overflow rate, m^3/day/m^2	13.44
(gpd/ft^2)	(330)
Weir overflow rate, m^3/day/m	46.38
(gpd/ft)	(3734)
Solids loading rate, kg MLSS/day/m^2	50.19
(lb MLSS/day/ft^2)	(10.28)

Sludge Characteristics

Although the secondary clarifiers were operated at lower loading rates than Phase 2, the concentration of the recycled activated sludge was less than Phase 2 (Table 12). The primary reason is felt to be the higher SVI, 285, for Phase 2. The mixed liquor solids were lower than Phase 2 because the RAS concentration was less and because activated sludge was wasted from the system. As a result of a higher MCRT for Phase 3, sludge production per unit of substrate removed decreased from Phase 2.

TABLE 12. PHASE 3 SLUDGE CHARACTERISTICS

MLSS, mg/l	3,726
MLVSS, mg/l	2,861
Return Activated Sludge TSS, mg/l	11,498
Return Activated Sludge Rate, % Q	
Rectangular clarifier	32.7
Circular clarifier	33.0
SVI, ml/gm	185
Mean Cell Residence Time, days	84
Sludge Production	
mass VSS produced/mass BOD removed	0.220
mass VSS produced/mass T-COD removed	0.111
mass VSS produced/mass S-COD removed	0.293

Oxygen Utilization and Nitrogren Conversion

Oxygen utilization efficiency was very low in Phase 3, 23.9 percent as shown in Table 13. The explanation of the low efficiency is the large quantity of oxygen lost through leaks in the tent, meaning that a large amount of oxygen added to the tent was simply to maintain pressure under the tent.

Oxygen consumption per pound of substrate removed was less in Phase 3 than in Phases 1 and 2, in spite of a longer MCRT than Phase 2. The explanation is felt to be the smaller mass of microbes under aeration, and hence there would be less oxygen used for endogenous respiration on a daily basis.

In contrast to Phases 1 and 2, incomplete ammonia oxidation occurred. Since MCRT was more than sufficient for growth of nitrifying organisms, it is probable that these organisms were killed in the final clarifier due to low D.O. conditions in the sludge blanket. Since no D.O. measurements were made on the sludge blanket contents, no proof of this hypothesis exists.

TABLE 13. PHASE 3 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen supplied kg O_2 supplied/m^3 (lb O_2 supplied/10^6 gal)	0.368 (3069)
mass O_2 supplied/mass BOD removed	4.35
mass O_2 supplied/mass T-COD removed	2.76
Oxygen Consumption kg O_2 consumed/m^3 (lb O_2 consumed/10^6 gal)	0.088 (732)
mass O_2 consumed/mass BOD removed	0.90
mass O_2 consumed/mass T-COD removed	0.57
Oxygen Utilization Efficiency, %	23.9
Average Aeration Basin DO, mg/l	7.1
Ammonia Oxidation, %	84.6
Rate of Nitrification mass NH_3-N oxidized/day/mass MLVSS	0.020
Alkalinity Consumption mass alkalinity consumed/mass NH_3-N oxidized	6.44

PHASE 4

Background

Phase 4 began on April 4, 1973 and was terminated on April 24, 1973, a period of 21 days. The principal change from Phase 3 was an increase in the flow from approximately 3785 to 4920 m^3/day (1 to 1.3 mgd). The reduced volume aeration basin first used in Phase 3 was again used as were the circular and rectangular secondary clarifiers. Aeration detention time in Phase 4 decreased to 2.39 hrs. The overall objective of Phase 4 was to study system performance at higher hydraulic and substrate loadings than Phase 3.

Operational Problems

The principal operational problem of significance was the recurring problem of holes in the tent. The windy spring weather not only helped create new leaks, but caused some previously sealed leaks to be reopened. The large number of leaks resulted in a decreased oxygen concentration in the atmosphere and a resultant low mixed liquor dissolved oxygen concentration.

Plugging on the dissolved oxygen sampling apparatus at sampling point one, near the head of the aeration basin was another consistent problem. As a result of this plugging, no usable dissolved oxygen data was obtained for sample point one.

Aeration Basin-Secondary Clarifier Efficiency

Substrate removals in Phase 4 (Table 14) dropped slightly from the efficiencies experienced in the previous three phases. BOD removal exceeded 92 percent, but T-COD and S-COD removal efficiencies averaged only 84.5 and 71.4 percent, respectively. It is felt that the character of the sludge was responsible for these decreases in substrate and solids removal efficiencies. That the sludge character had changed is quite evident from the SVI, which averaged 240. The most likely explanation for this change in sludge character is the low aeration basin D.O. levels which were measured. This hypothesis is clouded somewhat because no D.O. concentrations were recorded at the head end of the aeration basin, due to the previously discussed operational problems with sampling apparatus.

TABLE 14. PHASE 4 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	114.1	9.0	92.1
T-COD, mg/l	238.1	37.0	84.5
S-COD, mg/l	99.3	28.4	71.4
Suspended solids, mg/l	52.2	5.2	90.0
Volatile suspended solids, mg/l	47.3	4.1	91.3
NH_3-N, mg/l	14.9	4.7	68.5
NO_3-N, mg/l		5.9	
Total TKN, mg/l	15.7	7.0	55.4
Alkalinity, mg/l	237.1	180.9	23.7
Turbidity, JTU		2.1	
pH, units	7.43	7.23	

Ammonia reduction averaged a low 68.5 percent, the lowest level observed in any of the six phases of study. It is very probable that nitrification was inhibited by the low dissolved oxygen concentrations measured in the aeration basin and possibly also in the secondary clarifier. Although dissolved oxygen measurements were not made in the secondary clarifiers, it it reasonable to presume that there was less than in the aeration basin effluent, which averaged 1.58 mg/l.

F/M loadings increased (Table 15) as expected, due mainly to an increase in flow, but also because primary effluent BOD was somewhat greater and because MLVSS concentration was intentionally decreased. The substrate removal rate continued to increase in response to the higher substrate loading.

TABLE 15. PHASE 4 LOADING PARAMETERS

F/M Ratios	
mass BOD applied/day/mass MLVSS	0.332
mass T-COD applied/day/mass MLVSS	0.689
mass S-COD applied/day/mass MLVSS	0.286
Volumetric Loadings	
kg BOD applied/day/m^3	0.819
(lb BOD applied/day/1000 ft^3)	(51.11)
kg T-COD applied/day/m^3	1.709
(lb T-COD applied/day/1000 ft^3)	(106.7)
kg S-COD applied/day/m^3	0.712
(lb S-COD applied/day/1000 ft^3)	(44.46)
Substrate Removals	
mass BOD removed/day/mass MLVSS	0.304
mass T-COD removed/day/mass MLVSS	0.583
mass S-COD removed/day/mass MLVSS	0.206
Rectangular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	11.00
(gpd/ft^2)	(270)
Weir overflow rate, m^3/day/m	60.42
(gpd/ft)	(4865)
Solids loading rate, kg MLSS/day/m^2	33.64
(lb MLSS/day/ft^2)	(6.89)
Circular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	16.78
(gpd/ft^2)	(412)
Weir overflow rate, m^3/day/m	57.84
(gpd/ft)	(4657)
Solids loading rate, kg MLSS/day/m^2	51.46
(lb MLSS/day/ft^2)	(10.54)

Flow to the circular secondary clarifier was greater than in Phase 3, due to both a greater system flow and a greater percentage of flow going to the circular clarifier. The circular clarifier surface overflow rate was still relatively low, 2.38 m^3/day/m^2 (412 gpd/ft^2). Less flow was diverted to the rectangular secondary clarifier than in Phase 3. Thus, for the rectangular clarifier, the surface and weir overflow rates and solids loading rates decreased from Phase 3.

Sludge Characteristics

As discussed, MLSS and MLVSS were intentionally decreased to increase the F/M loadings in the aeration basin. This decrease was accomplished by increasing the rate of sludge wasting. The RAS concentration also decreased as a result of the increased SVI. The RAS concentration averaged 7,541 mg/l and was 80.6 percent volatile solids (Table 16).

TABLE 16. PHASE 4 SLUDGE CHARACTERISTICS

MLSS, mg/l	3066
MLVSS, mg/l	2480
Return Activated Sludge TSS, mg/l	7541
Return Activated Sludge Rate, % Q	
Rectangular clarifier	40.3
Circular clarifier	39.3
SVI, ml/gm	240
Mean Cell Residence Time, days	42.1
Sludge Production	
mass VSS produced/mass BOD removed	0.230
mass VSS produced/mass T-COD removed	0.121
mass VSS produced/mass S-COD removed	0.341

The MCRT decreased as a result of the increased rate of wasting activated sludge. MCRT still remained relatively high, averaging 42 days.

As a result of the increased wasting rate, excess sludge production increased, but only slightly from Phase 3. Excess sludge production was still relatively low, however, averaging 0.231 pounds of VSS produced per pound of BOD removed. This low rate is attributed to the long MCRT and the endogenous nature of the sludge at such a high MCRT.

Oxygen Utilization and Nitrogen Conversion

Oxygen utilization efficiency continued to be low, averaging only 24.2 percent (Table 17). This low efficiency was due principally to the large quantity of oxygen lost through leaks in the tent. The aeration basin dissolved oxygen concentration averaged only 1.52 mg/l. Again, this is principally a result of leaks in the tent and a concommitant decrease in the oxygen concentration in the atmosphere under the tent.

TABLE 17. PHASE 4 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen Supplied	
kg O_2 supplied/m^3	0.246
(lb O_2 supplied/10^6 gal)	(2052)
mass O_2 supplied/mass BOD removed	2.85
mass O_2 supplied/mass T-COD removed	1.77
Oxygen Consumption	
kg O_2 consumed/m^3	0.072
(lb O_2 consumed/10^6 gal)	(605)
mass O_2 consumed/mass BOD removed	0.58
mass O_2 consumed/mass T-COD removed	0.36
Oxygen Utilization Efficiency, %	24.2
Average Aeration Basin DO, mg/l	1.52
Ammonia Oxidation, %	68.5
Rate of Nitrification	
mass NH_3-N oxidized/day/mass MLVSS	0.0297
Alkalinity Consumption	
mass alkalinity consumed/mass NH_3-N oxidized	5.48

It is important to note that a D.O. concentration of zero was often measured near the middle of the aeration basin. Although D.O. measurements were not made at the head of the aeration basin, it logically follows from the above that zero D.O. concentrations also occurred at the head of the aeration basin.

Oxygen consumed per mass of BOD removed, after accounting for oxygen required for ammonia oxidation, averaged only 0.25, the lowest of any phase studied. This low figure is not felt to be valid, and probably results from discrepancies in measuring the quantity of atmosphere lost through leaks in the tent.

The percentage of ammonia converted to nitrate averaged only 68.5 percent. This low percentage, as discussed previously, is believed to be a result of low aeration basin D.O., and a probable low D.O. concentration in the secondary clarifier. Low D.O. concentrations such as this probably resulted in the death of many nitrifying organisms, especially of the genus Nitrosomonas.

Conclusions

The purpose of Phase 4 was to illustrate the effect of increased F/M loadings upon the system performance at a decreased mean cell residence time. These increased loadings were accomplished by increasing the flow and decreasing the MLVSS, while MCRT was decreased by increasing the rate of wasting activated sludge.

The results obtained indicated a decrease in overall percentage removed of most constituents measured. After an analysis of all parameters, however, it is believed the decreased system performance was due to a low mixed liquor D.O., averaging only 1.52 mg/l for sample points two and three. It is important to note that zero D.O. was often observed at sample point two, and although D.O. could not be measured at sample point one due to inoperative apparatus, it can be presumed that zero D.O. conditions also occurred at the head end of the aeration basin. The low D.O. conditions are attributed to a low atmosphere oxygen concentration resulting from the bleed-in of large quantities of air to compensate for leaks in the tent. The low mixed liquor D.O. is felt to also be the cause of the high SVI and the low percentage removal of ammonia, both the result of a probable change in the character of the sludge.

PHASE 5

Background

Phase 5 began on April 25, 1973 and continued until May 14, 1973, a period of 20 days. The major changes from Phase 4 were an increase in flow to approximately 4564 m^3/day (1.47 mgd) and a decrease in MLVSS. As a result of the increased flow, the aeration basin detention time decreased to 2.14 hrs. The reduced volume aeration basin and both the circular and rectangular clarifiers continued in use from Phase 4.

The objective of Phase 5 was to continue study of system performance at higher hydraulic and substrate loadings than used in prior phases.

Operational Problems

No new operational problems were encountered in Phase 5. However, leaks in the tent continued to appear and plugging of the D.O. sampling devices continued to occur.

Aeration Basin-Secondary Clarifier Efficiency

Substrate removal efficiencies in Phase 5 (Table 18) were equal to or greater than those observed in Phase 4. BOD and T-COD removal efficiencies were essentially the same as in Phase 4, being 91.8 percent and 85.1 percent, respectively. S-COD removal averaged 81 percent substantially better than in Phase 4. Suspended and volatile suspended solids removals were good, averaging 93.5 percent and 92.2 percent, respectively.

TABLE 18. PHASE 5 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	109.7	9.0	91.8
T-COD, mg/l	276.0	41.0	85.1
S-COD, mg/l	174.6	33.2	81.0
Suspended solids, mg/l	69.4	4.5	93.5
Volatile suspended solids, mg/l	62.8	4.9	92.2
NH_3-N, mg/l	16.4	3.6	78.0
NO_3-N, mg/l		6.7	
Total TKN, mg/l	15.9	5.1	67.9
Alkalinity, mg/l	237.8	170.7	28.2
Turbidity, JTU		2.0	
pH, units	7.42	7.06	

Oxidation of ammonia was greater than in Phase 4, averaging 78 percent. It is interesting to note that 12.8 mg/l of ammonia was removed, but that the maximum nitrate concentration increase was only 6.7 mg/l. Nitrate production may have been slightly less than this, however, but must be expressed as a maximum since primary effluent nitrate was not measured.

pH decreased due to destruction of alkalinity nitrification but only by 0.36 units. This decrease is substantially less than in other phases, except for Phase 4, where the pH decrease was only 0.2 units.

F/M loadings increased (Table 19) in response to the combination of increased flow and the decreased MLVSS concentration. On a BOD basis, the F/M loading was 0.385. The substrate removal rate also increased, averaging 0.366.

Both secondary clarifiers were more highly loaded than in Phase 4, but loadings were still relatively conservative. For rectangular secondary clarifiers, surface overflow and weir overflow rates were 12.1 $m^3/day/m^2$ (297 gpd/ft^2) and 66.57 $m^3/day/m$ (5368 gpd/ft), respectively. For the circular secondary clarifier, these values were 19.2 $m^3/day/m^2$ (471 gpd/ft^2) and 66.05 $m^3/day/m$ (5318 gpd/ft), respectively.

Sludge Characteristics

MLSS and MLVSS were intentionally lower than in any prior phase, averaging 2,712 and 2,277 mg/l, respectively (Table 20). The lower solids concentrations were achieved by increasing the rate of sludge wasting. The concentration of the RAS was lower than in Phase 4, although the SVI decreased. This is attributed to an increase in the rate of sludge pumping from the rectangular secondary clarifier, and a consequent dilution of the RAS. The volatile fraction of the RAS increased slightly from Phase 4 and averaged 83.9 percent.

TABLE 19. PHASE 5 LOADING PARAMETERS

F/M Ratios	
mass BOD applied/day/mass MLVSS	0.385
mass T-COD applied/day/mass MLVSS	0.960
mass S-COD applied/day/mass MLVSS	0.664
Volumetric Loadings	
kg BOD applied/day/m^3	0.877
(lb BOD applied/day/1000 ft^3)	(54.73)
kg T-COD applied/day/m^3	2.192
(lb T-COD applied/day/1000 ft^3)	(136.8)
kg S-COD applied/day/m^3	1.434
(lb S-COD applied/day/1000 ft^3)	(89.50)
Substrate Removals	
mass BOD removed/day/mass MLVSS	0.366
mass T-COD removed/day/mass MLVSS	0.816
mass S-COD removed/day/mass MLVSS	0.530
Rectangular Clarifier Loading	
Surface overflow rate, m^3/day/m^2	12.10
(gpd/ft^2)	(297)
Weir overflow rate, m^3/day/m	66.67
(gpd/ft)	(5368)
Solids loading rate, kg M.SS/day/m^2	32.71
(lb MLSS/day/ft^2)	(6.70)
Circular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	19.19
(gpd/ft^2)	(471)
Weir overflow rate, m^3/day/m	66.05
(gpd/ft)	(5318)
Solids loading rate, kg MLSS/day/m^2	51.75
(lb MLSS/day/ft^2)	(10.60)

TABLE 20. PHASE 5 SLUDGE CHARACTERISTICS

MLSS, mg/l	2712
MLVSS, mg/l	2277
Return Activated Sludge TSS, mg/l	6932
Return Activated Sludge Rate, % Q	
Rectangular clarifier	42.2
Circular clarifier	37.9
SVI, ml/ gm	220
Mean Cell Residence Time, days	22.4
Sludge Production	
mass VSS produced/mass BOD removed	0.396
mass VSS produced/mass T-COD removed	0.162
mass VSS produced/mass S-COD removed	0.371

As a result of increased sludge wasting, the mean cell residence time decreased to 22.2 days. This MCRT is approximately one-half the MCRT of Phase 4. Excess sludge production increased to 0.333. This is in response to an increased rate of sludge wasting, which lowered the MCRT and decreased the loss of cell mass by endogenous respiration. This excess production is still quite low compared to conventional activated sludge however, indicating that substantial cell mass is still being destroyed by endogenous respiration.

Oxygen Utilization and Nitrogen Conversion

As in Phase 4, oxygen utilization efficiency remained very low, averaging only 26.4 percent (Table 21). Again, this is attributed to the large quantity of atmosphere and thus oxygen, lost through leaks in the tent, aeration basin D.O. increased only slightly to 1.69 mg/l. This low value is the result of a low oxygen concentration in the atmosphere and the rapid rate of utilization at the F/M loading utilized.

Although D.O. overaged 1.0, 1.9 and 2.1 mg/l at the head end, middle, and effluent end of the aeration basin respectively, zero D.O. conditions also occurred at the head end, but measurements were only made at the head end on two days, and on these days zero D.O. conditions were not noted at the middle or effluent end of the basin.

Oxygen consumption still remained at a very low level, 0.37 after accounting for oxygen required for nitrification. This is not believed to be

a valid figure and is felt to be influenced by unexplainable discrepancies in correcting for leaks in the tent.

TABLE 21. PHASE 5 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen supplied	
kg O_2 supplied/m^3	0.309
(lb O_2 supplied/10^6 gal)	(2583)
mass O_2 supplied/mass BOD removed	3.81
mass O_2 supplied/mass T-COD removed	1.49
Oxygen consumption	
kg O_2 consumed/m^3	0.082
(lb O_2 consumed/10^6 gal)	(684)
mass O_2 consumed/mass BOD removed	0.97
mass O_2 consumed/mass T-COD removed	0.38
Oxygen Utilization Efficiency, %	26.4
Average Aeration Basin D.O., mg/l	1.69
Ammonia Oxidation, %	78.0
Rate of Nitrification	
mass NH_3-N oxidized/day/mass MLVSS	0.0470
Alkalinity Consumption	
mass alkalinity consumed/mass NH_3-N oxidized	5.22

As discussed, only 78 percent of the ammonia was oxidized. Inhibition of nitrification reactions is felt to have been caused by the occasional periods of zero D.O. in the aeration basin as well as occasional periods of zero D.O. in the secondary clarifiers. Again, alkalinity consumption in the nitrification reaction was less than stochiometric requirements, averaging 5.44 pounds of alkalinity per pound of ammonia oxidized.

PHASE 6

Background

Phase 6 began the day following termination of Phase 5, May 15, 1973 and continued for 119 days until September 10, 1973. In addition to increasing flow rate to 7064.4 m^3/day (1.84 mgd), MLVSS was also decreased to 2,121 mg/l. At this increased flow rate, detention time was a relativey low 1.7 hrs. The aeration basin and secondary clarifiers were identical to Phase 5.

The objective of Phase 6 was to determine what effluent quality could be produced at a relatively low aeration detention time. Another objective was to try to increase mixed liquor D.O. by feeding more pure oxygen to raise the percentage of oxygen in the atmosphere under the tent.

Operational Problems

The RAS pump on the rectangular clarifier caused problems on several occasions in Phase 6. On several occasions, the pump was down for service for up to a day. Near the end of Phase 6, the pump was found to have a cracked impeller and it was replaced.

Dumping of septic tank pumping trucks caused a degree of operational difficulty because the feed rate of pure oxygen had to be increased during periods they were dumping. It is probable that these shock loads of high strength septic wastes temporarily decreased the efficiency of treatment.

An extensive program of leak detection and patching was conducted throughout Phase 6. Unfortunately, this did not cut down on the leakage rate, and the worst leakage rates of the entire study occurred during the August and September of this phase. This was attributed to the extremely poor condition of the tent after a year and a half exposure to the elements.

Bulking in the rectangular secondary clarifier was noted on several occasions. This bulking was believed to have been caused by density currents. No action could be taken to eliminate these currents.

On several occasions, delivery of oxygen to the site was late and the system had to be run without pure oxygen feed for periods of several hours.

Aeration Basin-Secondary Clarifier Efficiency

Substrate removal efficiencies decreased to the lowest levels of any phase except Phase 2 (Table 22). BOD removal efficiency was only 89.4 percent, while T-COD and S-COD removals were 79.0 percent and 68.0 percent, respectively. Suspended and volatile suspended solids removal efficiencies were 86.6 percent and 87.2 percent, respectively.

Ammonia removal, 83.1 percent, was better than in Phases 4 and 5. This is probably a result of maintaining more consistent D.O. levels in the aeration basin as a result high feed rates of pure oxygen. At times, however, zero D.O. conditions continued to be observed, but the frequency of occurrence of these conditions was less than in Phases 4 and 5.

pH decrease averaged 0.77 units, the largest decrease noted except for Phase 1. The variation in pH decreases between phases cannot be explained, but probably is a function of buffering capacity not directly related to alkalinity of the wastewater.

F/M loadings reached their maximum levels during Phase 6 (Table 23). On a BOD basis, F/M loading averaged 0.454. This represents an increase of over six times the F/M loading used in Phase 1, with a corresponding decrease

in BOD removal efficiency from 98 percent to 89.4 percent. On a T-COD basis, however, F/M loading in Phase 6 was over 21 times that in Phase 1, yet the percentage T-COD removal was greater in Phase 6. Essentially the same relationships existed for substrate removal rates between Phases 1 and 6.

TABLE 22. PHASE 6 PRIMARY AND SECONDARY EFFLUENT CHARACTERISTICS

	Primary Effluent	Secondary Effluent	Percent Change
BOD, mg/l	95.3	10.2	89.3
T-COD, mg/l	239.4	50.3	79.0
S-COD, mg/l	100.0	32.0	68.0
Suspended solids, mg/l	44.1	5.9	86.6
Volative suspended solids, mg/l	37.6	4.8	87.2
NH_3-N, mg/l	18.3	3.1	83.1
NO_3-N, mg/l		8.1	
Total TKN, mg/l	19.7	6.1	69.0
Alkalinity, mg/l	221.7	150.9	31.9
Turbidity, JTU		2.5	
pH, units	7.4	6.63	

Secondary clarifier loading conditions remained very conservative even at the maximum hydraulic loading. Surface overflow rates were 16.38 and 22.41 $m^3/day/m^2$ (402 and 550 gpd/ft^2) for the rectangular and circular clarifiers, respectively. Weir overflow rates were 90.2 and 77.2 $m^3/day/m$ (7260 and 6214 gpd/ft) for the rectangular and circular clarifiers, respectively.

The sludge characteristics and the oxygen utilization and nitrogen conversion data are shown in Table 24 and Table 25, respectively.

TABLE 23. PHASE 6 LOADING PARAMETERS

F/M Ratios	
mass BOD applied/day/mass MLVSS	0.454
mass T-COD applied/day/mass MLVSS	1.137
mass S-COD applied/day/mass MLVSS	0.478
Volumetric Loadings	
kg BOD applied/day/m^3	0.996
(lb BOD applied/day/1000 ft^3)	(60.31)
kg T-COD applied/day/m^3	2.424
(lb T-COD applied/day/1000 ft^3)	(151.3)
kg S-COD applied/day/m^3	1.013
(lb S-COD applied/day/1000 ft^3)	(63.23)
Substrate Removals	
mass BOD removed/day/mass MLVSS	0.409
mass T-COD removed/day/mass MLVSS	0.898
mass S-COD removed/day/mass MLVSS	0.325
Rectangular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	13.97
(gpd/ft^2)	(343)
Weir overflow rate, m^3/day/m	76.94
(gpd/ft)	(6195)
Solids loading rate, kg MLSS/day/m^2	52.34
(lb MLSS/day/ft^2	(10.72)
Circular Clarifier Loadings	
Surface overflow rate, m^3/day/m^2	13.44
(gpd/ft^2)	(330)
Weir overflow rate, m^3/day/m	46.38
(gpd/ft)	(3734)
Solids loading rate, kg MLSS/day/m^2	50.19
(lb MLSS/day/ft^2)	(10.28)

TABLE 24. PHASE 6 SLUDGE CHARACTERISTICS

MLSS, mg/l	2491
MLVSS, mg/l	2121
Return Activated Sludge TSS, mg/l	8360
Return Activated Sludge Rate, % Q	
Rectangular clarifier	37.9
Circular clarifier	41.3
SVI, ml/gm	117
Mean Cell Residence Time, days	32
Sludge Production	
mass VSS produced/mass BOD removed	0.269
mass VSS produced/mass T-COD removed	0.122
mass VSS produced/mass S-COD removed	0.331

TABLE 25. PHASE 6 OXYGEN UTILIZATION AND NITROGEN CONVERSION

Oxygen Supplied	
kg O_2 supplied/m^3	0.430
(lb O_2 supplied/10^6 gal)	(3586)
mass O_2 supplied/mass BOD removed	4.93
mass O_2 supplied/mass T-COD removed	2.42
Oxygen Consumption	
kg O_2 consumed/m^3	0.208
(lb O_2 consumed/10^6 gal)	(1734)
mass O_2 consumed/mass BOD removed	2.38
mass O_2 consumed/mass T-COD removed	1.17
Oxygen Utilization Efficiency, %	48.4
Average Aeration Basin D.O., mg/l	1.2
Ammonia Oxidation, %	83.1
Rate of Nitrification	
mass NH_3-N oxidized/day/mass MLVSS	0.060
Alkalinity Consumption	
mass alkalinity consumed/mass NH_3-N oxidized	5.73

SECTION 6

DISCUSSION

NITROGEN INTERACTIONS

Conversion of ammonia to nitrates was essentially 100 percent in the first two phases of the study, but thereafter varied within the range of 68 to 86 percent. This indicates that the growth of nitrifying bacteria was somewhat inhibited in the latter four phases. It is believed that this inhibition of nitrification resulted from a combination of:

1. low average D.O. conditions in the aeration basin in the latter three phases,

2. transient zero D.O. conditions in the aeration basin in the latter three phases,

3. sludge blanket depths between 1.52 and 2.13 m (5 and 7 ft) in the secondary clarifiers in the latter four phases, and

4. successively shorter detention times with each phase which may have allowed short circuiting in the aeration basin during peak flow periods.

The influence of low averaged D.O. conditions upon nitrification in the study is questionable, since the lowest average D.O. concentration was 1.2 mg/l. It is generally believed that inhibition of nitrifying bacteria will not occur at D.O. concentrations greater than 1.0 mg/l. However, the occurrence of transient zero D.O. conditions in the aeration basin during peak flow periods definitely affected nitrification. Since D.O. measurements were not made in the secondary clarifier(s) on a continuous basis, it is not known with certainty whether zero D.O. conditions developed in the sludge blanket of the secondary clarifiers. Comparison of the nitrate concentration in the aeration basin with the concentration in the secondary effluent, however, points strongly to the existence of denitrification in the sludge blanket, and thus zero D.O. in the sludge blanket during the latter three phases. Table 26 compares nitrate and ammonia changes through the secondary clarifier.

This comparison illustrates several important facts. First, the ammonia concentration decreased in the secondary clarifier in each of the six phases, indicating some nitrification in the clarifier. This is also demonstrated by an increase in nitrate in the secondary clarifier in Phases 1, 2, and 3. Secondly, denitrification occurred in the clarifiers in Phases 4, 5, and 6. This is demonstrated by a nitrate decrease in the secondary clarifiers. The

summation of these two reactions is equal to the net nitrogen change, as shown in Table 26. In Phases 1 and 2, there was essentially no change within the limits of normal error. In Phases 3 to 6, it appears that nitrification occurred in the secondary clarifier, followed by denitrification, resulting in a net nitrogen loss (as nitrogen gas) from the system. The demarkation between the nitrification and denitrification reactions most likely corresponds with depletion of D.O. in the sludge blanket.

TABLE 26. NITROGEN CHANGES IN THE SECONDARY CLARIFIERS

	Phase 1	Phase 2	Phase 3	Phase 4	Phase 5	Phase 6
Nitrate						
Aeration Basin	15.70	13.50	7.85	6.61	7.71	9.10
Secondary Clarifier	16.20	13.70	8.00	5.90	6.70	8.10
Change	+0.50	+0.20	+0.15	-0.71	-1.01	-1.00
Ammonia						
Aeration Basin	0.80	0.16	2.25	5.30	3.79	3.40
Secondary Clarifier	0.40	0.08	1.60	4.70	3.60	3.10
Change	-0.40	-0.08	-0.65	-0.60	-0.19	-0.30
Net Nitrogen Change	+0.10	+0.12	-0.40	-1.31	-1.20	-1.30

NOTE: All units are in mg/l

A possible correction for this condition in the secondary clarifiers could have been an increase in D.O. in the aeration basin effluent. However, when this was tried on several occasions, turbidity of the secondary effluent would increase and overall effluent quality would deteriorate. In addition to this correction, return activated sludge pumping rate was also increased. The increased pumping rates did not alleviate the situation.

Also of interest is the net change in nitrogen content between the primary effluent and the secondary effluent. In the primary and secondary effluent, nitrogen is present as total TKN and nitrate. Nitrogen losses from the system include nitrogen gas produced by denitrification and nitrogen in waste activated sludge. Table 27 was prepared to compare the estimated amount of nitrogen wasted from the system as sludge with actual nitrogen lost as sludge from the system. Phase 1 is not shown in the table because of a lack of total TKN data and Phase 2 is not included because primary effluent nitrate concentrations were not measured, and it is felt that they were greater than zero. For all other Phases, primary effluent nitrate was assumed to be zero.

Although the results of the two calculations don't compare exactly, a general idea of nitrogen losses in WAS is obtained. The difference between the two methods is the result of subtracting many numbers of the same general magnitude, as well as using a cellular composition and assuming that it remained constant with time.

TABLE 27. NITROGEN LOSSES

	Phase 3	Phase 4	Phase 5	Phase 6
Mass TKN-N removed, kg/day (lb/day)	(106.6)	(94.7)	(127.1)	(208.9)
Mass NO_3-N produced, kg/day (lb/day)	(67.7)	(64.2)	(81.9)	(124.5)
Mass N lost by denitrification, kg/day (lb/day)	(11.1)	(14.3)	(14.7)	(20.0)
Estimated mass N in WAS, kg/day (lb/day)	(27.8)	(16.2)	(30.5)	(64.4)
Mass VSS in WAS, kg/day (lb/day)	(115.3)	(219.0)	(404.5)	(270.2)
Mass N in WAS, kg/day (lb/day)	(14.2)	(26.9)	(49.8)	(33.3)

SLUDGE SETTLEABILITY

The best laboratory measure of mixed liquor settleability is the SVI test. A variety of activated sludge conditions affect sludge settleability. Principal among these factors are:

1. volatility of the sludge,
2. nutrient balance in the primary effluent,
3. formation of nitrogen gas in the sludge due to denitrification, and
4. F/M loading rate in the aeration basin.

Sludge volatility is not believed to have been a factor in this study because it remained within a fairly well defined range throughout. Volatility of the mixed liquor solids varied from 77 percent to 85 percent. Neither was nutrient balance in the primary effluent felt to be a factor in the study. The Tapia facility receives sewage of primarily domestic origin which is properly balanced with nutrients for the growth of organisms in the activated sludge process.

Denitrification in the secondary clarifier did have an effect upon sludge settelability in this study, as previously shown by the denitrification which occurred in Phases 4, 5, and 6. Whether denitrification occurred within 30 min, the time required for SVI testing, cannot be stated.

F/M loading affects settling by influencing what growth phase the microorganisms are in at the effluent end of the aeration basin. At low F/M loadings, microorganisms are in the endogenous or death growth phase and settle very poorly because of their colloidal nature, low density, and destruction of the slime layer. At high F/M ratios, microorganisms are in the log growth phase and contain large quantities of bound water, making them approximately the same density as the liquid fraction of the mixed liquor. Protozoans and rotifers are present in small numbers, while free swimming bacteria are comprising the majority of the sludge. Poor settling is generally observed because of the large quantity of free-swimming bacteria. It is generally recognized that there is an optimum F/M loading between these two growth phase conditions.

A plot of SVI versus F/M loading (Figure 7) does not indicate the existence of an optimum F/M loading for this study. A curve is shown in Figure 7, however, to depict what is felt to define conditions for the study. The curve was drawn in an effort to place less emphasis on data where SVI is felt to have been influenced by denitrification. Based upon this curve, optimum F/M loadings appear to be above 0.5 mass BOD/day/mass MLVSS.

In addition to the effects of the loading characteristics on the highest SVI's, it is felt that the low rate of sludge wasting had an additional effect on the SVI. During the periods of time where sludge volume index was higher than 200, a sample of the sludge was microscopically examined. This examination revealed the fact that there was a considerable amount of trash dispersed

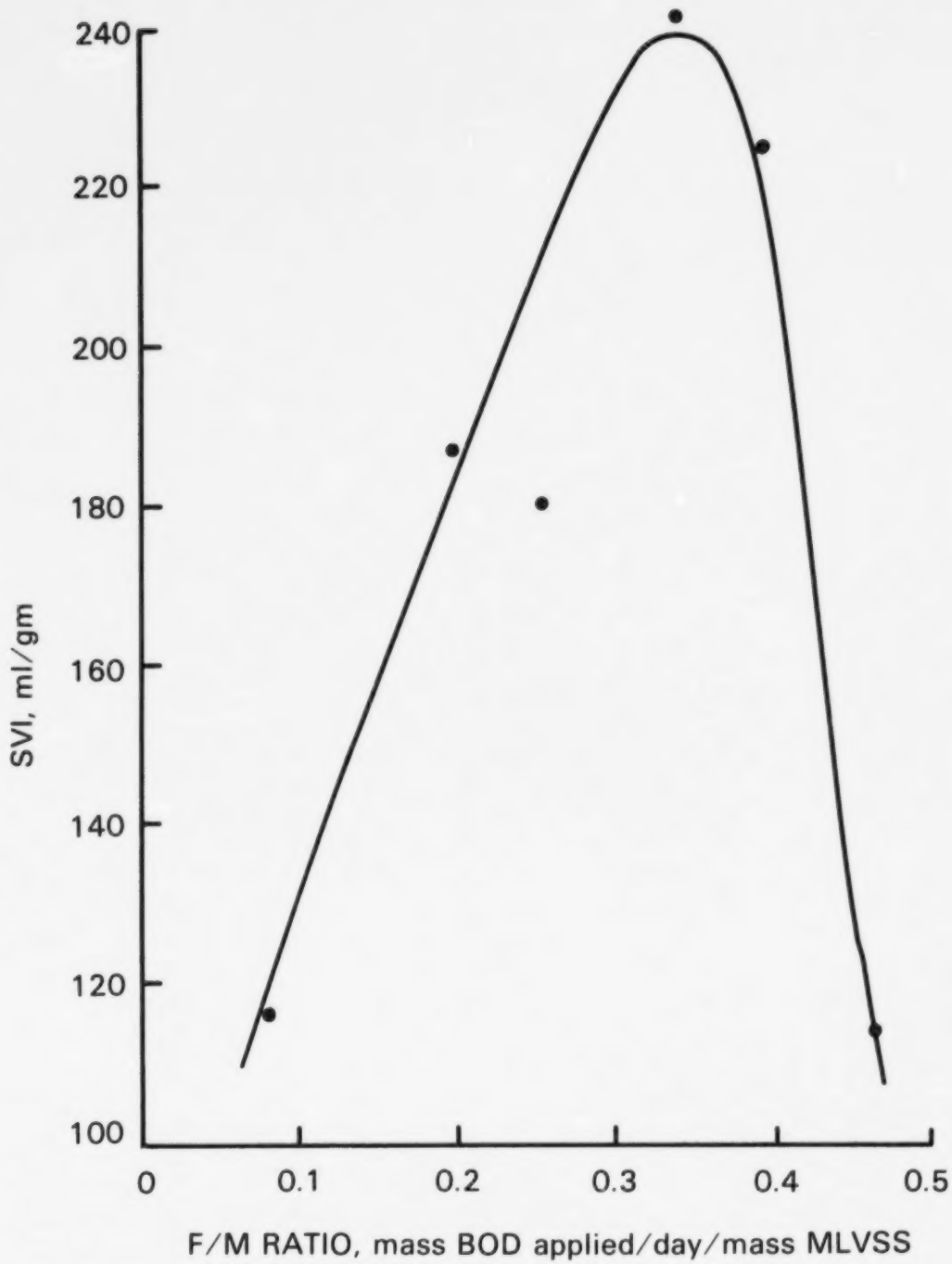

Figure 7. SVI versus F/M loading.

throughout the sludge. The trash consisted of grease particles and inert materials. This trash buildup seemed to coincide with the instances where SVI's were considerably high and wasting rates were low. At least three of these time periods of trash buildup are illustrated on Figure 7 as the highest SVI points on the graph.

SVI values obtained in this study are higher than those considered to be "good" in activated sludge processes using air aeration and are considerably higher than those found in other oxygenation studies. However, it should be realized that SVI is only an indicator of process performance, and the matter of ultimate importance is effluent quality. In this study, average secondary effluent BOD (before chlorination) varied between phases from 2 mg/l to 10 mg/l, and suspended solids from 4.8 mg/l to 13 mg/l. This effluent quality would be considered truly superb for most plants in the United States. However, the only acceptable effluent quality from the Tapia Water Reclamation Facility is an effluent with a continuous BOD concentration of less than 5 mg/l and a suspended solids concentration of 2 mg/l. Because high SVI values were obtained, this cannot be taken to mean that the process was ineffective, but only that SVI is a relative parameter and not directly related to plant effluent quality.

OXYGEN REQUIREMENTS

Provision of oxygen in the activated sludge process is necessary for cellular synthesis and endogenous respiration. Oxygen consumption in the process is correlatable with the quantity of substrate removed. Theoretically, one pound of oxygen is required for the biological stabilization of one pound of carbonaceous BOD, or slightly more if the microorganisms are maintained in the endogenous growth phase.

At long mean cell residence times, net oxygen requirements per quality of substrate removed increased for two reasons. First, the growth of nitrifying bacteria occurs (assuming other conditions are favorable), requiring oxygen at a theoretical rate of 4.57 pounds per pound of ammonia removed. Secondly, a high MCRT allows a longer time for cellular endogenous respiration with a resultant increase in oxygen consumption.

In this study, oxygen consumption was calculated by an oxygen balance, according to the formula: Consumption = (pure oxygen added) + (oxygen in bleed-in air) - (oxygen lost by leaks) - (oxygen in aeration basin effluent).

Results of the oxygen consumption calculations are presented in Table 28. These figures indicated definite descrepancies in either the measurements or the assumptions made in the calculations.

This is rather obvious for Phases 2-5 in which, on the average, less than one pound of O_2 was consumed per pound of BOD_5 removed, even though substantial endogenous respiration and nitrification were occurring. Since the calculated estimates for these four phases are in such apparent error, it is questionable whether the estimates from Phases 1 and 6 can be considered correct either.

The problem of leaks in the tent compounded the problem of maintaining D.O. in the aeration basin. A portion of the overall problem was also caused

by the location of adding bleed-in air to the system. Bleed-in air was added on the suction side of the recirculation blower, which in turn decreased the concentration of oxygen in the recirculation stream. Table 28 also presents average values for air bleed in, recirculation, and percent oxygen in the atmosphere, and calculated percent oxygen in the recirculation stream.

TABLE 28. OXYGEN DATA

	Phase 1	Phase 2	Phase 3	Phase 4	Phase 5	Phase 6
Mass O_2 consumed/mass BOD removed	3.50	1.08	0.90	0.58	0.97	2.38
Total recirculation, m^3/min (cfm)	18.32 (647.0)	18.44 (651.0)	8.51 (300.5)	6.64 (234.5)	5.96 (210.3)	6.54 (231.0)
Bleed-in air, m^3/min (cfm)	0.21 (7.5)	0.48 (17.0)	1.46 (51.6)	1.64 (57.9)	1.74 (61.4)	1.61 (56.7)
O_2 in atmosphere, %	44.0	49.5	39.5	38.8	41.1	54.7
O_2 in recirculation, %	43.7	48.8	36.3	34.4	35.2	46.2
Gas input, m^3/m^3 (ft^3/gal)	7.48 (1.0)	3.81 (0.51)	3.74 (0.50)	2.39 (0.32)	2.02 (0.27)	1.65 (0.22)

The table indicates that bleed-in did not significantly lower the recirculation oxygen content in Phases 1 and 2, but in the latter four phases lowered the recirculation oxygen content by 10 to 15 percent. This decrease is significant because it decreased the driving force for transfer of oxgyen to the mixed liquor. This was one factor tending to decrease lower mixed liquor D.O. concentrations than desired.

A second factor promoting low aeration basin D.O. concentration was the relatively small amount of recirculation required to maintain a "mixed" condition in the aeration basin. Table 28 presents results of calculations for gas phase inputs (recirculation plus pure oxygen) per gallon of wastewater. Thus, as wastewater flow increased, the quantity of gas (pure oxygen plus recirculation) decreased significantly. This decrease, by over a factor of four, allowed less gaseous-liquid phase contact and, therefore, less transfer of oxygen to the mixed liquor.

These two factors, dilution of recirculation with bleed-in and decreased gas phase input per gallon of wastewater, significantly affected the ability to maintain sufficient mixed liquor D.O. Unfortunately, the second of these factors was becoming worse at a time when the F/M loading on the system was being steadily increased.

EXCESS SLUDGE PRODUCTION

Excess sludge produced in the activated sludge process is the summation of waste activated sludge, skimmings from the secondary clarifiers, and suspended solids in the secondary effluent. The quantity of excess sludge produced is primarily a function of the length of time cell material remains in the system. As mean cell residence time increases, the growth curve of the microorganisms proceeds further and further into the endogneous phase of growth. A natural consequence is loss of cell mass by auto-oxidation.

Figure 8 presents a plot of pounds of excess VSS produced per pound of BOD removed versus MCRT. Each point represents the average of data for a one month period, except for Phase 1, which is represented by one point. As expected, excess VSS production decreased as MCRT increased.

More importantly, however, is the small quantity of solids produced per pound of BOD removed. Values in this study were about 40 percent less than in a conventional activated sludge process. Decreased sludge production is not due to "magical" properties of the Simplox process itself, but only that aeration with high purity oxygen allows more solids to be carried in the system, and thus allows operation at longer mean cell residence times. As Figure 8 shows, operation at these long MCRT results in significant decreased sludge production.

ALKALINITY CONSUMPTION

In the biological conversion of ammonia to nitrite, and thence nitrate, natural alkalinity is consumed in addition to ammonia alkalinity. The following equations represent biochemical reactions consuming alkalinity in the nitrification process:

$$2NH_4HCO_3 + 4O_2 \text{----} 2HNO_3 + 4H_2O + 2CO_2$$

$$2HNO_3 + C_A(HCO_3)_2 \text{----} C_A(NO_3)_2 + 2CO_2 + 2H_2O$$

$$2NH_4HCO_3 + 4O_2 + C_A(HCO_3)_2 \text{----} C_A(NO_3)_2 + 4CO_2 + 6H_2O$$

On a stochiometric basis, 7.18 pounds of alkalinity are destroyed for each pound of ammonia oxidized.

During this study, two phases had alkalinity reduction greater than stochiometric requirements, while four had less. Table 29 presents calculated values for alkalinity consumption.

OPERATION CONSIDERATIONS

As was pointed out in an earlier section of this report, one of the objectives of this study was to ascertain whether or not savings could be achieved in capital costs due to the supposedly short durations of aeration time required for proper treatment. This objective is interpreted by the Las

Virgenes Municipal Water District as being a reduction in aeration tank size requirements for future expansions of the Tapia Water Reclamation Facility. The reduction of aeration tank capacity has to be of such magnitude to offset any additional operation and maintenance costs that may be experienced through the use of an enriched oxygen system. At this point the reader of this report should fully understand the basis upon which the District makes their interpretation. On a nationwide basis, it has generally been accepted that the design parameter for the aeration tank capacity has generally been in the area of 6 to 8 hrs of aeration time. To the Las Virgenes Municipal Water District this 6 to 8 hrs of aeration time is felt to be excessive. Consequently, in the design and operation of the Tapia Water Reclamation Facility, the air activated sludge aeration system was designed for 3.6 hrs of aeration time. It has been proven at the Tapia facility that this 3.6 hrs is more than adequate for not only carbonaceous removal but, additionally, for complete nitrification. Over and above this, the air activated sludge system at the Tapia facility has performed superbly at aeration times as low as 2 hrs. The guideline or the limiting factor for the operation at the Tapia facility is the effluent quality. The only acceptable effluent quality produced at Tapia is an effluent with a BOD concentration of 5 mg/l or less and a suspended solids concentration of 5 mg/l or less, a turbidity of one Jackson unit or less, and a total coliform index of less than 2.2/100 ml.

TABLE 29. ALKALINITY CONSUMPTION

Phase	Mass Alkalinity Consumed/ Mass NH_3 Oxidized
1	7.70
2	9.14
3	6.44
4	5.48
5	5.22
6	5.73

In view of the above information it is evident that in order for the Simplox oxygenation system to become a viable alternative for treatment at the Tapia facility, not only must the aeration time be extremely low but at the same time the oxygen system must produce a superb quality effluent.

Throughout the entire study, effluent quality remained foremost as a limiting parameter. That is, operational changes were made to continually produce the best effluent possible.

From an operational control standpoint the oxygenation system did not require significant changes in normal operating procedures related to an activated sludge system. However, operator attention to these procedures appeared to have a more intense requirement than the normal air system. Table 30 depicts a comparison of costs associated with the Simplox oxygen system versus the Tapia air system. Operation labor costs for the oxygen

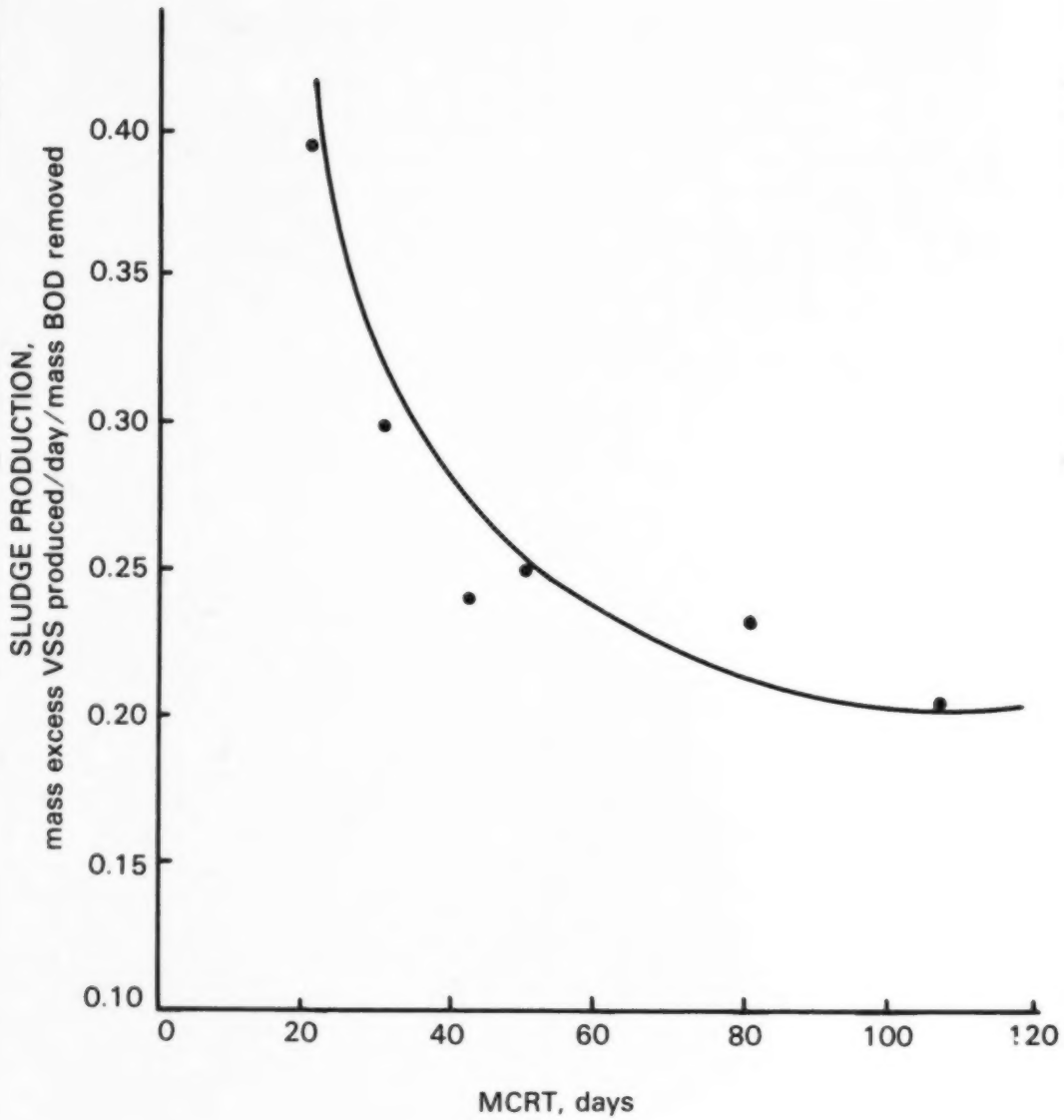

Figure 8. Sludge production versus MCRT.

system as shown in Table 30 depict this increase. It was found that more operator time had to be expended on the liquid oxygen control, conversion and distribution system than would be required on a normal air compressor for an air activated sludge system. However, it should be pointed out that in the $31.00/mil gal on the O_2 system, considerable labor was expended for the constant repair of tent leakage. Laboratory control of the oxygen system appears to be about the same as what is required in an air activated sludge system. Power, as shown in Table 30, is predominately the power requirement to operate the mixing aeration blower on the oxygen system. As expected, of course, this is considerably lower than the power requirement for the operation of air compressors to support the air activated sludge system. Oxygen costs as depicted on Table 30 include the cost of delivery of liquid oxygen to the Tapia site.

TABLE 30. OPERATION COST COMPARISON

	Simplox System	Tapia Air System
Operation labor, $/1000 m^3	8.19	6.34
Laboratory, $/1000 m^3	6.34	6.08
Power, $/1000 m^3	3.17	6.34
Oxygen, $/1000 m^3	12.68	0
Total, $/1000 m^3	30.38	18.76

NOTE: $/1000 m^3 x 3.785 = $/$10^6$ gal

The four categories that are used for the cost comparison in Table 30 are felt to be the categories that have a direct effect on the total cost in comparing the two systems. Items such as chemical costs for chlorine used for disinfection are equal to both systems, therefore were not used. In addition, the operation labor category for both systems includes the incidental preventative maintenance on equipment such as adjustments of packing on pumps, bearing inspections, etc.

Although the data illustrated that there was some reduction in the amount of excess sludge produced in the oxygen system, it was not felt to be large enough to warrant reduction in aerobic digester sizing at the Tapia Facility. Therefore no credit was given to the oxygen system for reduction in sludge handling costs.

From the costs shown on Table 30, it is apparent that the increased oepration costs for the oxygen system over the air system is predominately the purchased oxygen. Admittedly, amounts of oxygen were lost unnecessarily through the tent leakage throughout the study. However, if the oxygen costs were removed entirely, the related costs are still near the same. Further, since the Tapia air system can produce an acceptable effluent at an aeration time of only 2.0 hrs as compared to the oxygen system's 1.7 hrs, the difference of 0.3 hr for aeration time could not justify a large enough savings in capital costs to offset the total operation and maintenance costs of the oxygen system.

Additionally, at the 1.7 hrs aeration time, the oxygen system cannot continually produce the effluent quality needed at the Tapia facility.

CALCULATION PROCEDURES

Because of a lack of continuity by various researchers in the definitions of several important parameters, it is felt that the definition of certain parameters should be presented at this point to eliminate any confusion in the interpretation of results. These key parameters were calculated according to the following definitions:

Mean Cell Residence Time (MCRT) - Mass of volitale suspended solids in the system (aeration basin, aeration basin effluent channel, secondary clarifier sludge blanket, and return activated sludge line) divided by the mass of volatile suspended solids wasted per day (volatile waste activated sludge, volatile skimmings, and volatile solids in the final effluent). Units-days.

Substrate Removal Rate - The mass of substrate (expressed as BOD, T-COD or S-COD) removed per day per mass of mixed liquor volatile suspended solids in the aeration basin. Units-substrate removed/day/mass MLVSS.

Food - Microorganism Loading (F/M Loading) - The mass of substrate (expressed as BOD, T-COD or S-COD) applied to the aeration basin per day per mass of mixed liquor volatile solids in the aeration basin. Units-mass substrate applied/day/mass MLVSS.

Aeration Detention Time - The average length of time wastewater remains in the aeration basin. Calculated by dividing aeration basin capacity by the summation of primary effluent and return activated sludge. Units-hours.

The definitions of other terms used throughout this report are presented in the glossary.